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Pollution Control Technical Manual for Exxon Donor Solvent Direct Coal Liquefaction

TRW, Inc., Redondo Beach, CA

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The document is one of six technical handbooks prepared by EPA to help government officials granting permits to build synfuels facilities, synfuels process developers, and other interested parties. They provide technical data on waste streams from synfuels facilities and technologies capable of controlling them. Process technolgies covered in the manuals include coal gasification, coal liquefaction by direct and indirect processing, and the extraction of oil from shale. The manuals offer no regulatory guidance, allowing the industry flexibility in deciding how best to comply with environmental regulations.

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FOREWORD

The purpose of the Pollution Control Technical Manuals (PCTMs) is to provide process, discharge, and pollution control data in summarized form for the use of permit writers, developers, and other interested parties. The PCTM series covers a range of alternate fuel sources, including coal gasification and coal liquefaction by direct and indirect processing, and the extraction of oil from shale.

The series consists of a set of technical volumes directed at production facilities based upon specific conversion processes. The entire series is supplemented by a pollution control technology appendix volume which describes the operation and application of approximately 50 control processes.

All PCTMs are prepared on a base plant concept (coal gasification and liquefaction) or developer proposed designs (oil shale) which may not fully reflect plants to be built in the future. The PCTMs present examples of control applications, both as individual process units and as integrated control trains. These examples are taken in part from applicable permit applications and, therefore, are reflective of specific plants. None of the examples are intended to convey an Agency endorsement or recommendation but rather are presented for illustrative purposes. The selection of control technologies for application to specific plants is the exclusive function of the designers and permitters who have the flexibility to utilize the lowest cost and/or most effective approaches. It is hoped that readers will be able to relate their waste streams and controls to those presented in these manuals to enable them to better understand the extent to which various technologies may control specific waste streams and utilize the information in making control technology selections for their specific needs.

The reader should be aware that the PCTMs contain no legally binding requirements or guidance, and that nothing contained in the PCTMs relieves a facility from compliance with existing or future environmental regulations or permit requirements.

Herbert L. Wiser
Acting Deputy Assistant Administrator
Office of Research and Development
U.S. Environmental Protection Agency

ABSTRACT

The Environmental Protection Agency (EPA), Office of Research and Development, has undertaken an extensive study to determine synthetic fuel plant waste stream characteristics and pollution control systems. The purpose of this and all other Pollution Control Technical Manuals (PCTMs) is to convey this information in a manner that is readily useful to designers, permit writers, and the public.

The Exxon Donor Solvent (EDS) direct coal liquefaction PCTM addresses the coal liquefaction technology being developed by Exxon Research and Engineering Company. Two configurations of the EDS process are considered in detail in this manual. These are the base case configuration and the Market Flexibility Sensitivity (MFS) configuration. The PCTM also considers a single feed coal (Illinois No. 6 bituminous), although the effects of other coal types on emissions and controls are described. Throughout the PCTM, the flow rates of waste streams and process streams presented correspond to an EDS commercial plant processing 1,134 Mg/hr (30,000 tons per stream day) of "as received" Illinois No. 6 coal in the liquefaction area. This EDS plant would produce about 9,580 m^3 (60,240 barrels) fuel oil equivalent per stream day of liquid products for the base case design, or 11,300 m^3 (71,080 barrels) fuel oil equivalent per stream day of liquid and gas products for the MFS case design.

This manual describes the two configurations of the EDS process, characterizes the waste streams produced in each medium, and discusses the array of commercially available controls which can be applied to the base plant waste streams. From these generally characterized controls, example control trains are constructed for each medium, illustrating the function of some typical integrated control systems. Control and control system cost and performance estimates are presented, together with descriptions of the discharge streams and secondary waste streams.

A summary of environmental and control technology considerations is presented. This summary includes a listing of data limitations and needs.

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GLOSSARY OF ACRONYMS

ADA	Anthraquinone disulfonic acid
ADIP	Shell-patented acid gas removal process based on DIPA or MDEA
AGR	Acid gas removal
BFOE	Barrels fuel oil equivalent, 1 BFOE = 6,050,000 Btu (LHV)
BOD	Biochemical oxygen demand
BSRP	Beavon Sulfur Removal Process
COD	Chemical oxygen demand
CRA	Compression-refrigeration-absorption
CRF	Capital Recovery Factor
DEA	Diethanolamine
DIPA	Diisopropanolamine
DOE	Department of Energy
ECLP	Exxon Coal Liquefaction Pilot Plant
EDS	Exxon Donor Solvent
EGD	Effluent Guidelines Division, Office of Water Regulations and Standards, EPA
EP	Extraction Procedure
EPA	Environmental Protection Agency
EPRI	Electric Power Research Institute
ERDA	Energy Research and Development Administration
ER&E	Exxon Research and Engineering Company
ESP	Electrostatic precipitator
FBC	Fluidized bed combustion
FGD	Flue gas desulfurization
FGR	Flue gas recirculation
FGT	Flue gas treatment
HHV	High Heating Value
IERL	Industrial Environmental Research Laboratory
LEA	Low excess air
LHV	Lower Heating Value
LNB	Low NO_x burners
LPG	Liquified petroleum gas

GLOSSARY OF ACRONYMS (Continued)

MAF	Moisture and ash free
MEA	Monoethanolamine
MDEA	Methyldiethanolamine
MFS	Market Flexibility Sensitivity
NO_x	Nitrogen oxides
NMHC	Non-methane hydrocarbons
NPDES	National Pollution Discharge Elimination System
NSPS	New Source Performance Standards
OAQPS	Office of Air Quality Planning and Standards, EPA
OFA	Overfire Air Injection
OPTS	Office of Pesticides and Toxic Substances, EPA
OSW	Office of Solid Wastes, EPA
PCB	Polychlorinated Biphenyls
PCTM	Pollution Control Technical Manual
PNA	Polynuclear Aromatics
POM	Polycyclic organic matter
RCRA	Resource Conservation and Recovery Act
RL	Reduced Load
ROM	Run of Mine
SCOT	Shell Claus Off-Gas Treatment
SCR	Selective Catalytic Reduction System
SNG	Substitute Natural Gas
SNPA	Societe Nationale des Petroles d"Aquitaine
SO_x	Sulfur oxides
SRC	Solvent Refined Coal
TDS	Total dissolved solids
TEA	Triethanolamine
TGT	Tail gas treatment
TOC	Total organic carbon
TSP	Total suspended particulates
TSS	Total suspended solids

GLOSSARY OF ACRONYMS (Continued)

TVA	Tennessee Valley Authority
VGO	Vacuum gas oil
VOC	Volatile Organic Compounds
W-L	Wellman-Lord

CONVERSION FACTORS

1.0 kg [kilogram] = 2.205 lb [pound (mass)]

1.0 Mg [megagram (metric ton)] = 1.102 ton [short ton (2000 lb)]

1.0 m^3 [cubic meter] = 264.2 gal [gallon]

1.0 ℓ [litre] = 0.2642 gal [gallon]

1.0 Nm^3/hr [normal cubic meter (at 0°C) per hour] = 37.32 scfh [standard cubic feet = (at 60°F) per hour]

1.0 GJ [gigajoule] = 0.9479 x 10^6 Btu [British thermal unit]

1.0 MW [megawatt] = 3.413 x 10^6 Btu/hr [British thermal unit per hour]

1.0 MJ/s [megajoule per second] = 3.413 x 10^6 Btu/hr [British thermal unit per hour]

1.0 kWh [kilowatt hour] = 3413 Btu [British thermal unit]

1.0 MJ/Nm^3 [megajoule per normal cubic meter at 0°C)] = 25.40 Btu/scf [Btu per standard cubic foot (at 60°F)]

1.0 g/Nm^3 [gram per normal cubic meter (at 0°C)] = 0.413 gr/scf [grains per standard cubic foot (at 6°C)]

1.0 MPa [megapascal] = 9.87 atmosphere

1.0 kmol = 22.4 Nm^3 (at 0°C and 1 atmosphere)

Prefixes

T = tera = 10^{12}

k = kilo = 10^3

G = giga = 10^9

m = milli = 10^{-3}

M = mega = 10^6

μ = micro = 10^{-6}

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1. INTRODUCTION

1.1 PURPOSE AND USE OF PCTM'S

It is envisioned that an environmentally acceptable synthetic fuels industry will be developed for future U.S. energy production. As part of this overall effort, the Environmental Protection Agency (EPA), Office of Research and Development, has for the past several years undertaken an extensive study to determine synthetic fuel plant waste stream characteristics and potentially applicable pollution control systems.

The purpose of the Pollution Control Technical Manuals (PCTM's) is to convey in a summarized and readily useful manner information on synfuel waste stream characteristics and pollution control technology as obtained from studies by EPA and others. The documents provide waste stream characterization data and describe a wide variety of pollution controls in terms of estimated performance, cost, and reliability. The PCTM's contain no legally binding requirements, no regulatory guidance, and include no preference for process technologies or controls. Nothing within these documents binds a facility to utilize a specific emission control process nor relieves a facility from compliance with existing or future environmental regulations or permits.

The Pollution Control Technical Manuals consist of several discrete documents. There are process-specific PCTM's and a more general appendix volume which describes over fifty pollution control technologies. Application of pollution controls to a particular synfuel process is described in each process-specific manual. The volumes currently contemplated are:

Pollution Control Technical Manual for Lurgi-Based Indirect Coal Liquefaction and SNG

Pollution Control Technical Manual for Koppers-Totzek-Based Indirect Coal Liquefaction

Pollution Control Technical Manual for Exxon Donor Solvent Direct Coal Liquefaction

Pollution Control Technical Manual for Lurgi Oil Shale Retorting with Open Pit Mining

Pollution Control Technical Manual for Modified In-Situ Oil Shale Retorting Combined with Lurgi Surface Retorting

Pollution Control Technical Manual for TOSCO II Oil Shale Retorting with Underground Mining

Control Technology Appendices for the Pollution Control Technical Manuals

By focusing on specific process technologies, the PCTM's attempt to be as definitive as possible on waste stream characteristics and control technology applications. This focus does not imply any EPA recommendations for particular process or control designs. Those described in the manuals are intended as representative examples of processes and control technologies that might be used. The organization of the PCTM's from process description through waste stream characterization and control technology evaluation provides the user with a number of alternative approaches to understanding the environmental consequences in operating synthetic fuel plants.

Control technology configurations presented in the PCTM's reflect pollutant removal levels which are believed to be achievable with currently available control technologies based upon existing data. Since there are no domestic commercial scale synfuels facilities, the data base supporting this document is from bench and pilot synfuel facilities, developers' estimates, engineering analyses, analogue domestic industries, and non-U.S. commercial synfuel plants. As commercial synthetic fuel plants are built, the EPA will continue conducting research in order to develop a more comprehensive data base. In the interim, the Agency encourages facility planners, permit officials, and other interested parties to take advantage of the information contained in these documents.

1.2 CONTENT OF THIS PCTM

1.2.1 Synfuels Process Addressed

This PCTM addresses the Exxon Donor Solvent (EDS) process for direct coal liquefaction. In direct liquefaction technologies such as EDS, coal is converted directly into liquids, as distinguished from indirect liquefaction processes where the coal is first gasified and the product gas then synthesized into liquid products. The direct liquefaction approach receiving the most attention involves hydrogenation of the coal at intermediate temperature (650-900°F, or 340-480°C) and elevated pressure (100-200 atm, or 10-20 MPa), using a heavy coal-derived organic solvent as a medium to aid in adding the hydrogen to the depolymerized coal molecules. This hydrogenation approach is employed by EDS and by other direct liquefaction processes in advanced stages of development, including Solvent Refined Coal (SRC) and H-Coal. These processes produce a range of products -- including naphtha (which can be refined into gasoline and other fuels), a range of fuel oils (for domestic heating, industrial/utility boilers and stationary turbines), light hydrocarbon gases and other by-products such as sulfur, ammonia and phenols.

Research and engineering on the EDS process began at Exxon in 1966. A two-year test program has recently been completed on a 250 ton coal/day pilot plant in Baytown, Texas. No demonstration-scale or commercial-scale EDS facilities are contemplated at the present time, although Exxon considers the technology technically ready for commercial application, and although plans are underway for large-scale plants employing some of the other direct liquefaction processes.

The EDS process is fundamentally similar to the SRC and H-Coal processes in many respects: all of the processes involve the addition of hydrogen to finely-crushed coal at elevated temperature and pressure in a medium of hydrogen-donating solvent, followed by separation of the product liquids and recycle of the recovered solvent back to coal hydrogenation reactors. There are some significant differences between the processes, also; one unique feature of the EDS process is the use of a separate catalytic reactor to add hydrogen back into the recovered solvent prior to the return of the solvent to the coal hydrogenation reactors.

The similarities among the direct liquefaction processes will naturally lead to similarities in the types of environmental discharges from commercial plants employing these different processes. There will be differences in exact discharge stream identities, in exact discharge flow rates and in exact discharge compositions, depending upon which direct liquefaction process is employed, and depending upon how the process is designed; but the types of commercial discharges will be similar for all processes and designs. For example, all direct liquefaction processes will have an acid gas stream requiring control; all will have a high-organics wastewater stream; all will have a solid waste stream consisting largely of coal ash. The control technologies applicable to these types of streams will generally be the same, irrespective of the direct liquefaction processes- although performance and costs of the controls will necessarily vary as discharge stream flow and composition varies from process to process.

This PCTM addresses the EDS process specifically, due to the availability of pilot plant environmental data and good commercial-scale conceptual plant designs for the EDS process. Detailed stream identifications, discharge rates, and stream compositions presented in this manual -- and the related estimated performance and costs of alternative control techniques -- are based upon specific conceptual commercial EDS plant designs. However, the information presented in this manual concerning the types of streams requiring control -- and concerning the types of control techniques applicable to these stream types -- should be of general use in considering SRC and H-Coal facilities as well as EDS, and in considering EDS plant designs in addition to the specific designs utilized in this manual.

This PCTM addresses the discharges and controls associated with a commercial-scale EDS direct liquefaction plant, beginning with the receipt of the feed coal and other raw materials, and proceeding through the separation and storage of the direct liquid products which would normally then be transferred to a conventional refinery for further refining/upgrading. All ancillary operations associated with the direct liquefaction plant (e.g., steam and power generation, feedwater treatment) are addressed. The manual does not address discharges/controls associated with: the ultimate refining/upgrading of the direct liquids into marketable products, the distribution of these products to ultimate users, or the ultimate use of the products.

1.2.2 Information Presented in PCTM

Section 1 of the manual presents an introduction to the overall approach and content of the manual. Section 2 is an overview of the EDS process. Section 3 presents the flow and composition for each discharge stream within a conceptual commercial EDS plant design, broken down by process area within the plant. For each discharge stream thus identified, the manual summarizes (in Section 4) a wide range of alternative individual control techniques that might be considered for the control of that stream, providing such information as: control process principle; development status; pollutant removal performance; reliability; factors influencing performance; secondary discharge streams resulting from the control process; and advantages/disadvantages. For many of these alternative control techniques, further details concerning the technique are presented in the separate volume entitled "Control Technology Appendices for PCTMs", and design considerations and costs for applying the technique to the EDS commercial design are discussed in Section 4. Where a stream would require several individual control techniques in series (for example, a wastewater control system with three or four or more techniques combined), Section 4 also presents some examples of possible integrated control systems, to illustrate possible ways in which individual controls might be combined. Finally, in Section 5, the manual summarizes data gaps and limitations (concerning e.g., stream composition or control technique performance), in order to illustrate the limitations of the data base.

The PCTM addresses both uncontrolled, or primary, discharge streams (i.e., those that would result from an uncontrolled EDS facility), and secondary discharge streams (those which result from the application of a control technique to the EDS plant). The term "uncontrolled discharge stream" refers to a stream prior to the application of any controls; as such, an "uncontrolled discharge stream" is generally not a true discharge stream at all, since, in a commercial EDS facility, control techniques would normally be applied.

1.3 APPROACH FOR DEVELOPING PCTM

1.3.1 Plant Design and Uncontrolled Discharge Estimates

The discharge stream identifications, flow rates and compositions presented in this manual are based upon two conceptual commercial-scale EDS plant designs prepared by Exxon, using Illinois coal (5,6,7,8). One of the two conceptual plant designs, identified as the "base case", is fed with 30,000 tons of coal per stream day, and has a product output (in the form of naphtha, fuel oil, and liquified petroleum gas) of 60,240 barrels per stream day (fuel oil equivalent). In the "base case" design, hydrogen for the coal liquefaction reaction is generated through steam reforming of the plant's by-product synthetic natural gas. The second related conceptual plant design, identified as the "market flexibility sensitivity" (or MFS) case, utilizes 30,000 tons of coal per stream day and generates 71,080 barrels per stream day (fuel oil equivalent) of liquid and gaseous products. The MFS design differs from the base case primarily in that the MFS case generates the hydrogen for liquefaction by gasifying about half of the heavy residual organics remaining after the liquefaction/separation steps, rather than by steam reforming the synthetic natural gas; this approach leaves the MFS case with the synthetic natural gas as a plant product, and results in an increased plant efficiency relative to the base case. Both the base and the MFS cases are addressed throughout the PCTM, in order to illustrate the effect that plant design can have on discharges and controls. In fact, Exxon has recently introduced a third design (the "bottoms recycle" case), designed to increase the yield of lighter liquid products; the bottoms recycle design was not completed in time for inclusion in the PCTM.

In utilizing the Exxon commercial plant designs, the PCTM has relied upon heat and material balances prepared by Exxon. In order to confirm the flow rate of uncontrolled discharges within the designs, EPA has, where warranted, conducted independent engineering evaluation of the Exxon mass balances. Likewise, in estimating the composition of the discharges, the Exxon design estimates have been bolstered using: environmental data which have become available since the Exxon designs were completed; thermodynamic calculations; and engineering evaluation.

The PCTM focuses in Section 3 on an "uncontrolled" EDS plant -- i.e., a plant with no environmental controls at all. The philosophy is that, once the uncontrolled plant is defined, then the array of possible alternative control techniques can be presented (in Section 4) for each uncontrolled discharge stream. Since the Exxon designs for the base and MFS plants incorporated environmental controls, these controls have been "subtracted out" for the purposes of the uncontrolled plant description in Section 3, and for the purposes of estimating uncontrolled plant capital and operating costs in Section 2. In cases where there might be some question concerning whether some technique is primarily intended as an environmental control measure -- or whether it is instead a part of the EDS process which would be utilized even in the absence of environmental requirements -- this question is addressed in the appropriate part of Section 3. For example, acid gas removal processes -- intended to remove reduced sulfur and other contaminants from EDS process gases -- are considered part of the EDS process; the sulfur recovery and other steps which are then used to treat the removed acid gases, however, are considered to be environmental control measures.

Discharges from future commercial EDS plants (or discharges presented in environmental permit applications associated with such plants) could differ from those presented in the PCTM. One reason for such differences would be the uncertainty involved in estimating emission/effluent compositions at this time; as discussed elsewhere, the environmental data are limited, the available experimental facilities are not always representative of commercial plants in terms of environmental discharges, and there is always necessarily some uncertainty in making engineering estimates of trace components in discharges from complex systems. Beyond this uncertainty, however, there are other reasons why there will be differences between the PCTM and subsequent EDS facilities. Discharges will vary depending upon a number of factors:

- Feedstock properties. For example, the PCTM (and the Exxon base and MFS designs) are based upon an Illinois bituminous coal. The EDS process has also been piloted on sub-bituminous coal and lignite; discharges from a commercial plant processing these coals will be somewhat different from a plant processing Illinois coal.
- EDS plant design. The detailed design of the plant can affect the identity of the discharge streams as well as their flows and compositions. The PCTM is based upon the base and MFS cases. But a third case exists now (bottoms recycle), and variations on these cases are possible in the future.
- EDS plant operating conditions. Along with plant design, plant operating conditions can be varied in order to adjust product slate and production rates. Changes in operating conditions can influence discharges.

The approach that has been employed in the PCTM has been to use the best estimates/data currently available, and the feedstock, design details and operating conditions employed in the Exxon base and MFS conceptual designs.

1.3.2 Control Technology Evaluation

For each "uncontrolled" EDS plant stream identified in Section 3, the manual presents in Section 4 a range of possible individual alternative control techniques that might be considered for that stream. Key information concerning each control technique (as generally applied to synfuels processes) is summarized in a tabular format (e.g., process principle, development status, general performance, reliability, secondary streams, and advantages/disadvantages). For many of the individual control techniques, further discussion is provided in Section 4 of the specific performance and costs associated with the application of the technique to each pertinent stream in the EDS plant designs, stream by stream. Also for many of the techniques, further details supporting the Section 4 presentation are provided in the separate appendix volume. Where a stream would require several individual control techniques combined in some manner, Section 4 also presents some examples of how the individual techniques might be combined into integrated control systems, including assessment of the performance and costs of these integrated control examples applied to the EDS designs.

The evaluation of individual control techniques in the manual is based upon literature information, contacts with control equipment vendors, information on experience with the control technique in related conventional industries, any experimental data regarding the performance of the technique on EDS or other synfuels streams, and engineering evaluation. In some cases, simple models have been developed providing technique cost as a function of major variables, such as capacity. Since essentially none of the control techniques have been operated routinely on a large EDS (or other synfuels) facility, some extrapolation has generally been required in projecting control performance on an EDS plant. Of particular difficulty has been the estimation of process reliability, capital and operating costs.

In the Section 4 summary of individual control techniques for each stream, an effort has been made to address essentially all of the possible alternative techniques that might be considered by a developer planning an EDS plant. However, the techniques discussed in detail in the Appendix -- and the techniques which are further discussed in Section 4, in terms of their application to the

EDS plant designs -- do not include the full list of techniques. Rather, these latter discussions address only those techniques which have received serious consideration to date from synfuels plant designers, or which offer potential for improved performance relative to commonly considered techniques.

Capital and operating costs for control techniques are presented on three bases, wherever possible: 1) cost per unit throughput; 2) absolute cost of the technique, as applied to the EDS plant designs; and 3) control technique cost as a percentage of the total cost of the uncontrolled EDS plant. In some cases, only the cost per unit throughput is shown for a technique; these cases result when the potential variability in the cost of the technique (e.g., due to possible variations in system design) is so great that the cost range would be too broad to be meaningful if expressed as absolute cost or as percentage cost. The cost estimates are considered accurate to an extent no greater than ± 50%.

As with the estimates of plant discharge flows/compositions, there might be differences between the performance and cost estimates presented in the PCTM for the control techniques, and the performances/costs that will be experienced in practice in future synfuels plants. One reason is the uncertainty in the estimates of flow rates and compositions of the streams being fed to the controls, as discussed in Section 1.3.1. In addition, there is uncertainty associated with estimating the performance and costs of a control, even if the feed stream to the control were defined exactly; design information to enable rigorous detailed design of many controls is not generally available; detailed designs are beyond the scope of this manual; and even detailed designs reflect some uncertainty. The possible effect on the control technique of some perhaps unanticipated, interfering trace contaminant in the EDS stream might not be predictable. Beyond the uncertainty in estimating control performance/costs for the EDS base and MFS designs, there is variability that can result from variations in these EDS design conditions in future, actual plants: variations in coal feedstock, plant design and plant operating conditions could significantly influence uncontrolled discharges, as discussed in Section 1.3.1, and thus influence control performance/costs.

Most significant of all, the performance and costs of a control technique can be affected tremendously by variations in the design of the control system in which the technique appears. For example, the performance and cost of one individual wastewater control technique, such as biological oxidation -- which is only one of several individual technques that might appear in a multi-component wastewater control system -- will depend upon: how the various individual wastewater streams are combined and fed into the various individual system components; which specific techniques precede the biox unit in series; and what organisms and operating procedures are selected for the biox unit. As another example, the performance/cost of flaring could vary significantly depending upon whether one central flarestack is used for the entire plant, or whether two separate smaller flares are employed. One purpose of the integrated control examples in Section 4 is to illustrate performance/costs for a few representative systems for which these fundamental control system design decisions have been made. A third example is that there is an "equipment quality - service factor" tradeoff which affects capital investment costs and operating costs. Some companies prefer to modify equipment or add additional equipment to a basic design to insure smoother, less troublesome operation. This results in a higher service factor and lower operating costs, but also increases capital investment cost. Other companies prefer to take the basic design, or minimize capital investment, with resulting lower service factor and higher operating costs. The tradeoff between maximizing service factor or minimizing capital investment is usually based on design philosophy and/or economics. The capital investment and operating costs presented in Section 4 will be subject to variation due to the "equipment quality - service factor" tradeoff.

Despite the unavoidable uncertainty/variability in any performance/cost estimates which are made at the present time, the performance and cost figures presented in this manual are felt to be effective relative indicators of control performance and costs. Throughout the PCTM, efforts have been made to quantify the uncertainty/variability where this is possible.

1.4 DATA BASE FOR PCTM

1.4.1 Data Base for Uncontrolled Discharge Estimates

The experimental data upon which the uncontrolled discharge estimates are based, result from two major sources:

1) The small pilot-scale data which were reflected in the Exxon conceptual commercial plant designs. Exxon's designs for the base case and MFS case EDS facilities reflect process and environmental data obtained by Exxon using a 1 ton coal/day small pilot-scale unit, and various smaller bench-scale experimental facilities. These designs were completed prior to the availability of significant data from the large, 250 ton coal/day Exxon Coal Liquefaction Pilot Plant (ECLP). Therefore, to the extent that the PCTM utilizes the discharge flow rates and compositions generated in the Exxon designs, the manual reflects Exxon's data from the 1 ton/day small pilot unit.

2) Source testing data from the 250 ton/day ECLP facility. EPA has completed a three-day environmental source sampling effort on the ECLP plant, and the data from this campaign have been utilized in updating and supplementing the discharge composition estimates from the earlier EDS designs. The EPA campaign included a broad screening effort addressing conventional pollutants, trace metals, and a wide range of organic compounds. In addition, Exxon has completed some environmental testing on the ECLP unit, and some of the Exxon data have been made available to EPA and have been utilized in the PCTM. Where appropriate, the ECLP data have been used in place of any corresponding design estimates (based on the 1 ton/day unit), or in place of calculated values.

In addition to the above major sources, data are available from three EPA source testing campaigns on a 30 ton coal/day Solvent Refined Coal pilot plant. These data, while not presented extensively in this manual, were helpful in confirming the data from comparable streams in the EDS plant.

There are some apparent limitations in this discharge stream data base. First, the substantive ECLP source testing was conducted at essentially one (representative) set of EDS conditions; thus, the data base does not cover a range of plant feedstock, operating or design conditions. Second, the plant is lacking certain discharge streams that would exist in a commercial facility; other streams that are present in ECLP might not be representative of the comparable commercial-scale stream. For example, ECLP does not include a bottoms processing operation (e.g., Flexicoking or gasification), so that the wastewater and waste ash streams associated with that operation are missing. ECLP does not include its own wastewater treatment plant, so that representative samples of wastewater treatment sludges cannot be obtained. Because the ECLP facility does not include the full array of valves, pumps, flanges, etc., at representative scale, EPA did not obtain data on fugitive organics emissions; nor was a representative flarestack available.

Additional environmental data from the ECLP facility are expected to become available in the future.

1.4.2 Data Base for Control Performance/Costs

There are essentially no data on the performance of control equipment operating continuously on an experimental EDS facility. The reason is that the experimental units do not include dedicated, representative control techniques; the units are built primarily to explore EDS process-related questions, not to demonstrate environmental control technology. In the ECLP facility, for example, recovered acid gases are directed to the sulfur recovery system in the refinery which adjoins the pilot plant, where the ECLP acid gases are mixed with gases from the refinery prior to treatment; thus, no data can be obtained on the performance of the sulfur recovery system on the ECLP acid gases alone. Likewise, ECLP wastewaters are directed to the neighboring refinery's wastewater treatment system. Solid wastes typical of a commercial EDS facility -- e.g., Flexicoker and gasifier ash, wastewater treatment sludges -- are not generated at ECLP, and hence are not treated/disposed. Some pilot plants for direct liquefaction processes other than EDS, do include dedicated wastewater treatment systems, but these systems are not considered to be representative, so the data can be used only with caution in this PCTM. A small wastewater treatment train has been installed on a slipstream of the wastewaters from the H-Coal pilot plant in Catlettsburg, Kentucky, but no results were available in time to be included in the PCTM.

As a result, the only real control performance data obtainable from EDS would be that developed by taking samples of streams in the EDS facilities, and treating them separately in remote experimental control facilities. In fact, Exxon had reportedly planned to treat ECLP wastewater samples in bench-scale wastewater treatment equipment, but the results from this effort were not available for inclusion in the PCTM. Similar types of testing are being conducted by others on samples of wastewaters from other direct liquefaction experimental facilities, for SRC and H-Coal; the available data are limited, but have been utilized where felt to be applicable to the EDS case.

1.5 HOW TO USE THE PCTM

There are two major components of the PCTM: 1) identification and characterization of uncontrolled discharge streams, in Section 3; and 2) assessment of a range of alternative control options for each uncontrolled stream thus identified, in Section 4.

Section 3 begins (Section 3.2) with a listing of all of the discharge streams (both uncontrolled and secondary) in the EDS base and MFS commercial plant designs. This list is broken down by process area within the plant and, within each process area, is further broken down by medium (gaseous, liquid, solid). The bulk of Section 3 which then follows (Section 3.3) provides a detailed description of each process area and of each discharge stream, giving flow and composition information for each stream. This detailed description follows the same order as the listing in Section 3.2, and is broken down by process area and, within process area, by medium, in the same manner. Furthermore, Section 3.3 includes index terms in the upper outside corner of each page, indicating the process area and the stream(s) covered on that page. Therefore, a user of the manual who is interested in only one medium or in only a few specific streams, can readily locate the streams of interest in the listing in Section 3.2; and, using the process area and stream number information obtained from the listing, the user can then quickly flip through Section 3.3, using the index terms, to find the detailed discussion of the particular streams of interest.

In Section 3.4, summary tables are presented which list uncontrolled emissions fro the total EDS base and MFS plants, broken down according to each of the uncontrolled streams discussed in Section 3.3.

At the end of Section 3 (Section 3.5), there are two cross-reference tables, linking Section 3 with Section 4 (and with the Appendix). In the first of these cross-reference tables, for each uncontrolled discharge stream identified in Section 3, the table indicates the precise subsection in Sections 3 and 4 where the user will find detailed stream characterization and discussion of the control techniques applicable to that stream. For completeness, the table includes not only the uncontrolled discharge streams discussed in Section 3, but also secondary discharge streams which are introduced in Section

4. The second cross-reference table is broken down by pollutant, rather than by stream, indicating where in Section 4 and the Appendix the user will find the control techniques applicable to each individual pollutant. The user interested in a few specific streams, or a few specific pollutants, can then readily locate the detailed discussions of the pertinent control techniques for those streams or those pollutants.

The control technique discussion in Section 4 is broken down by medium (Sections 4.2, 4.3, 4.4). Within each medium, the discussion is broken down by "source type"; a source type includes individual discharge streams which have similar control requirements. At the beginning of each source type subsection, a summary table is presented summarizing key information for each alternative control technique that might be considered for use on that source type. Then, for each source type, each specific stream within that source type is discussed, stream by stream. Both uncontrolled and secondary streams are addressed. In these stream-by-stream discussions, further details are provided for many of the individual control techniques, concerning the performance and costs associated with the application of each technique to each stream.

Where a stream would require several individual control techniques in some combination, the controls for that stream are divided into functional areas. For example, if a wastewater stream might utilize bulk organics removal and dissolved gas removal and further dissolved organics removal, then these steps would be defined as "functions", and the individual control techniques that fit into each of these functional areas would be identified. In general, functional areas would normally fit together in some fairly consistent sequence, and are usually defined such that either only one individual control technique, or no technique, would be selected from each function in assembling the total, combined system. With this functional method of presentation, the user then has the flexibility of considering the individual techniques in any desired combination that he chooses, knowing the sequence in which the functions normally fit together. For streams that do require multiple control techniques in combination, Section 4 also includes some representative examples of how the variety of individual techniques from different functions might be combined into specific integrated control systems.

Throughout Section 4, index terms appear in the upper outside corner of each page, indicating the medium, source type, stream number, and (where appropriate) functional area covered on that page. Referring to these index terms -- especially with the aid of the cross-reference tables in Section 3.5 -- the user should be able to readily locate the page on which the controls are discussed for any stream of interest.

The Appendix provides substantial additional details for many of the individual control techniques, beyond the level of detail in Section 4. The Appendix presents information general to the control technique, and does not address the application of the technique to the EDS plant designs specifically. For example, the Appendix might present curves showing the cost of a technique as a function of key variables. This information supports the presentation in Section 4, where, for example, the cost numbers emphasized would be the costs (taken from the Appendix curves) for applying the technique to the specific EDS plant designs, considering the values for the major variables associated with these designs.

In Section 5, a table is presented summarizing data gaps and limitations concerning the characterization of major discharge streams, and concerning the performance/cost of major control techniques. This table is broken down by stream, with the control techniques applicable to a given stream included under that stream.

2. PROCESS DESCRIPTION OVERVIEW

The EDS process is one of the more developed direct coal liquefaction processes. Direct coal liquefaction processes convert coal directly into liquid hydrocarbons whereas indirect coal liquefaction processes first convert coal into gaseous products (gasification step) and then catalytically react the gaseous products to form liquid hydrocarbons (liquefaction step) (1,2). The EDS process is fundamentally similar to other leading direct liquefaction processes in many aspects: all of the processes involve the addition of hydrogen to finely-crushed coal at elevated temperature and pressure in a medium of hydrogen-donating solvent, followed by separation of the product liquids and recycle of the recovered solvent back to the coal liquefaction reactors. There are also, of course, significant differences between the EDS and other direct liquefaction processes. One unique feature of the EDS process is the use of a separate catalytic reactor to add hydrogen back into the recovered solvent prior to recycling the solvent to the coal liquefaction reactors.

This section presents a brief discussion of the overall EDS process and identifies the major waste streams generated in the facility. The major variables that affect the emissions and the cost for an EDS plant without environmental controls will also be discussed.

2.1 OVERALL PROCESS DESCRIPTION

Two configurations for the EDS process are in advanced stages of development. These are the base case configuration and the Market Flexibility Sensitivity (MFS) case configuration. A third configuration (the bottoms recycle configuration) has been proposed by Exxon, but is covered only briefly in the PCTM because design details were not available in time for inclusion.

These alternative configurations, examined in this PCTM, are embodied in specific commercial plant conceptual designs prepared by Exxon. Where specific flows, conditions and compositions are referenced to the base, MFS, and bottom recycle cases in this manual, these are referring to the specific Exxon designs.

2.1.1 Base Case

Figure 2-1 presents a simplified flow diagram for the EDS plant based on the base case design. This design maximizes the production of C_4^+ liquids; light hydrocarbons (C_3^-) produced in the liquefaction reactor are steam reformed to generate hydrogen for in-plant use.

Raw coal is crushed to minus 8 mesh in the coal preparation area (at a rate of 27.2 Gg, or 30,000 tons/stream day in the Exxon commercial design). Crushed coal is then fed to the slurry drier, where it is mixed with recycled donor solvent to form a slurry; the slurry is dried to less than 4.0 wt % moisture based on dry coal feed. After drying, the coal slurry is pumped to reaction pressure, mixed with preheated hydrogen and heated in the slurry preheat furnace before entering the liquefaction reactor which operates at 700-750°K (800-890°F) and 13.9 MPa (2000 psig) (3). In the reactor, coal is liquefied in the presence of molecular hydrogen and hydrogenated donor solvent. The donor solvent facilitates the dissolution of coal, and donates hydrogen to the dissociated coal radicals. In the process, the donor solvent decomposes and becomes "spent".

The products from the liquefaction reactor are separated in a vapor-liquid separator and several distillation steps into a light hydrocarbon gas stream, spent donor solvent, a C_3 liquefied petroleum gas (LPG), a C_4 LPG, a naphtha (pentanes to 350°F normal boiling range), fuel oil, and a heavy vacuum bottom stream. The light hydrocarbon gas stream, scrubbed with diethanolamine (DEA) to remove NH_3, H_2S and CO_2, is then passed through a steam reforming unit where hydrogen for the liquefaction step is generated. The spent donor solvent is hydrogenated in a fixed bed catalytic reactor and is recycled back to the liquefaction reactor. The C_3 LPG, C_4 LPG, naphtha and fuel oil are stored as products (60,240 barrels fuel oil equivalent per stream day for the Exxon commercial design). The heavy vacuum bottoms are fed to a Flexicoking* unit where

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*"Flexicoking" is the service mark for a proprietary process developed by Exxon.

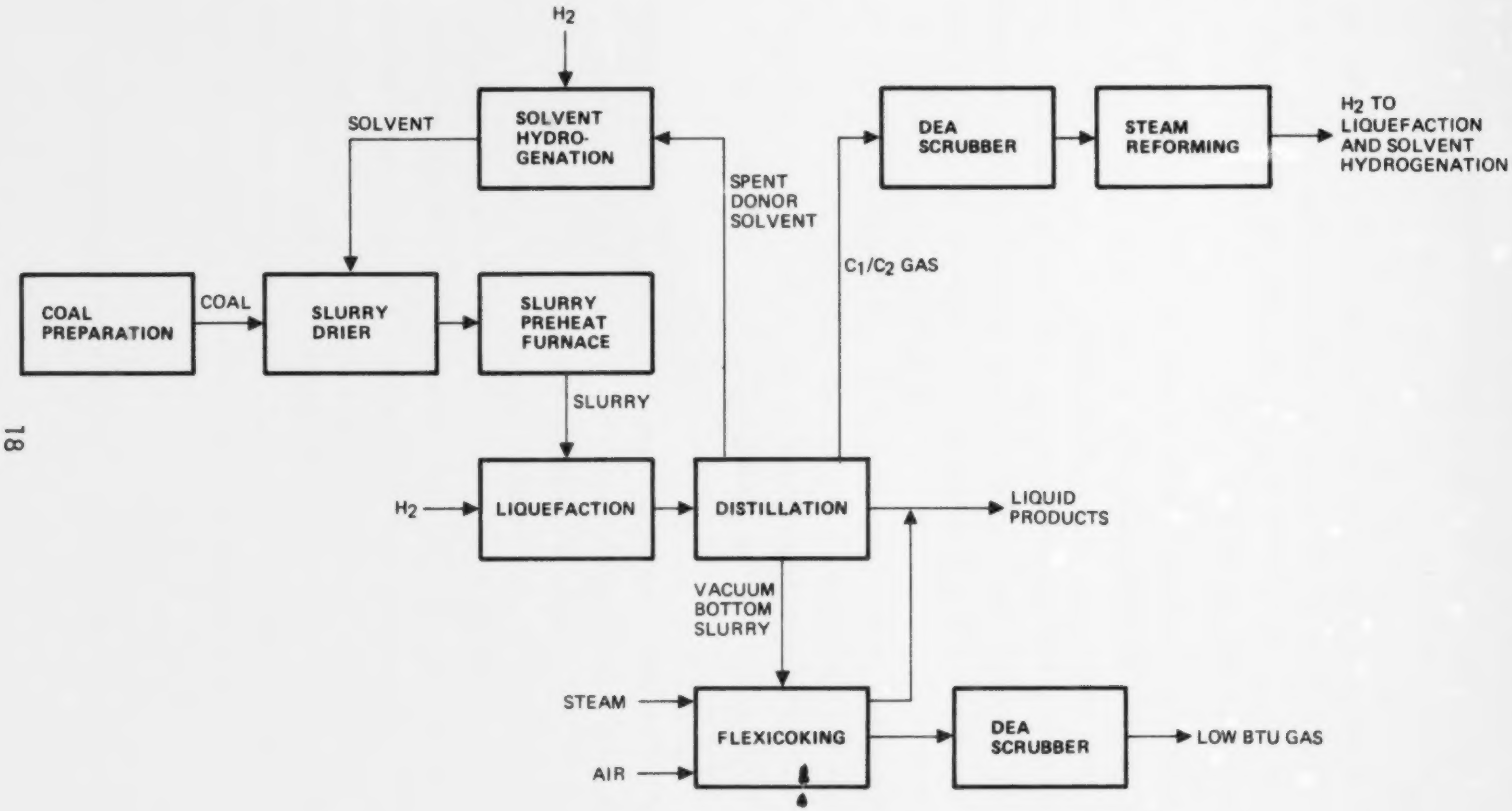
H2
SOLVENT
SOLVENT HYDRO-GENATION
SPENT DONOR SOLVENT
DEA SCRUBBER
STEAM REFORMING
H2 TO LIQUEFACTION AND SOLVENT HYDROGENATION
C1/C2 GAS
COAL PREPARATION
COAL
SLURRY DRIER
SLURRY PREHEAT FURNACE
SLURRY
H2
LIQUEFACTION
DISTILLATION
LIQUID PRODUCTS
VACUUM BOTTOM SLURRY
STEAM
AIR
FLEXICOKING
DEA SCRUBBER
LOW BTU GAS

Figure 2-1. Simplified Block Flow Diagram for Exxon Donor Solvent Process - Base Case

steam and air are added to produce additional distillate liquid products and a low Btu fuel gas for process furnaces. The vacuum bottoms are the organic- and ash-containing heavy residues remaining after rigorous vacuum distillation has removed essentially all of the desired liquid products. In the Flexicoking unit, essentially all organic material in the vacuum bottoms is recovered as liquid products or fuel gas. A small amount of residual carbon is rejected with the ash.

There are several major types of emissions from an uncontrolled EDS plant. These include acid gases, fugitive emissions, inorganic- and organic-laden wastewaters, and inorganic- and organic-based solid wastes. The acid gases are generated by acid gas removal systems (DEA) in the liquid distillation area and the Flexicoking and hydrogen generation area, and by steam stripping of sour water. These streams contain high concentrations of reduced sulfur and nitrogen species such as H_2S and NH_3. Fugitive emissions include particulate emissions resulting from coal storage and handling, and organics from leaks, product storage and material handling. Inorganic- and organic-laden wastewaters include process wastewaters and plant area runoff; process wastewaters result from unreacted steam used in liquefaction, from oxygen or moisture in the coal, and from water added to dissolved output slurry in order to prevent salt deposition. Many of these wastewater streams contain high concentrations of phenols and ammonia. Inorganic-based solid wastes include ash generated in the Flexicoking area and spent catalysts generated in various operations such as hydrotreating and sulfur recovery. Organic-based solid waste consists primarily of sludges from treatment of wastewater. In general, the characteristics of waste streams generated from the EDS process are similar to those generated from other direct liquefaction processes, in that the same major components are typically found in corresponding waste streams. The differences between EDS waste streams and other waste streams are primarily in waste generation rates, in concentration levels of major components, and in types and concentration levels of minor and trace components.

2.1.2 Market Flexibility Sensitivity (MFS) Case

Figure 2-2 is a simplified block flow diagram for the EDS plant based on the MFS case design. The major difference between the base case and the MFS case design is the approach used for the production of hydrogen. In the MFS plant, hydrogen is produced by partial oxidation of about 50% of the vacuum

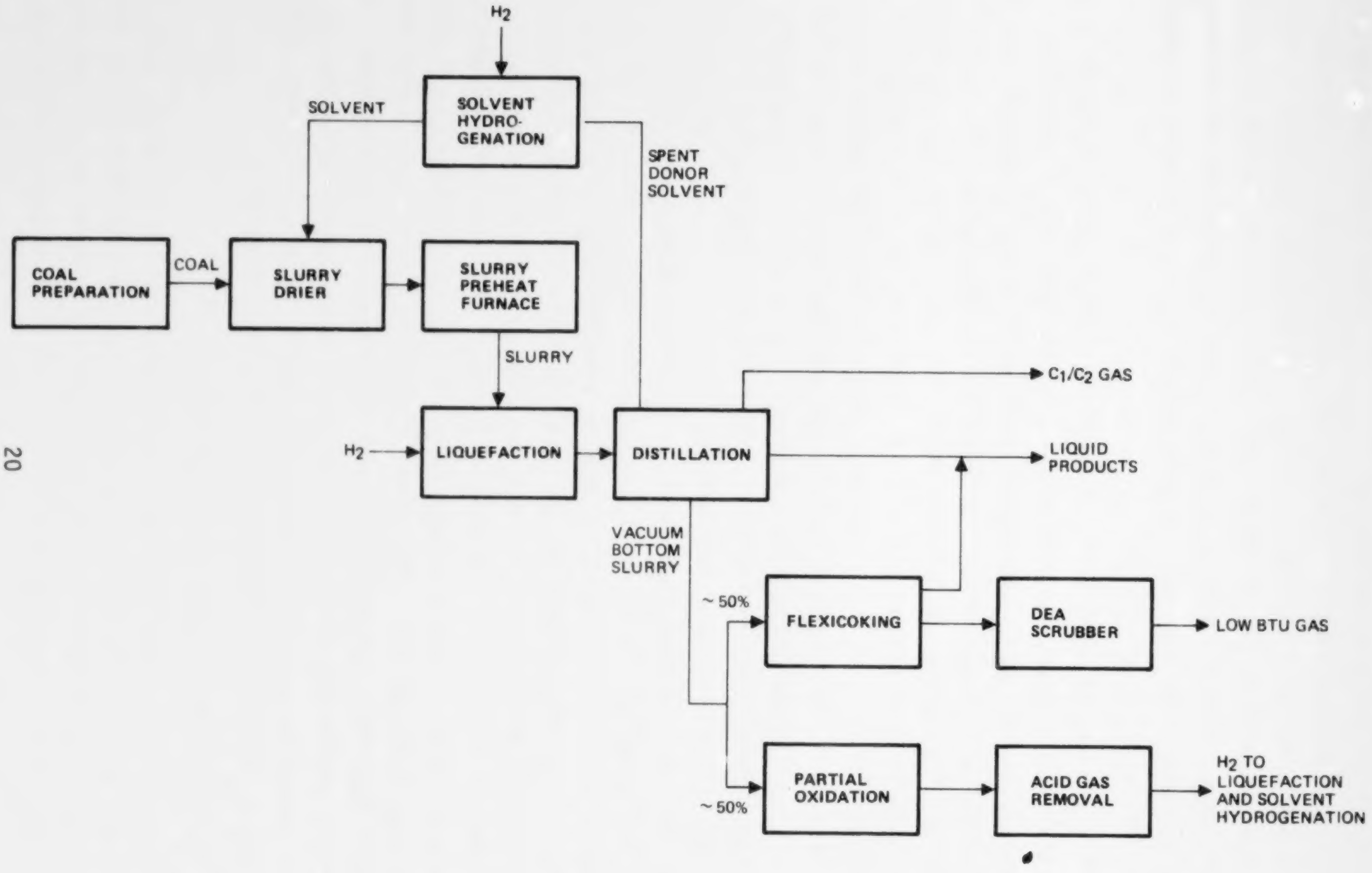
H_2
SOLVENT
SOLVENT HYDRO-GENATION
SPENT DONOR SOLVENT
COAL PREPARATION
COAL
SLURRY DRIER
SLURRY PREHEAT FURNACE
SLURRY
H_2
LIQUEFACTION
DISTILLATION
C_1/C_2 GAS
LIQUID PRODUCTS
VACUUM BOTTOM SLURRY
~ 50%
FLEXICOKING
DEA SCRUBBER
LOW BTU GAS
~ 50%
PARTIAL OXIDATION
ACID GAS REMOVAL
H_2 TO LIQUEFACTION AND SOLVENT HYDROGENATION

Figure 2-2. Simplified Block Flow Diagram for Exxon Donor Solvent Process -- MFS Case

bottoms instead of by steam reforming of light hydrocarbon gases as in the base case. The remaining 50% of the vacuum bottoms are fed to a Flexicoking unit to produce low Btu gas for plant fuel and additional liquid products. The light hydrocarbon gases recovered by eliminating the steam reforming step are sold as pipeline gas after undergoing further treatment. Total gaseous and liquid products from the Exxon commercial MFS design amount to 71,080 barrels fuel oil equivalent per stream day, for the same coal feed rate to the liquefaction section as in the base case design. Thus, the differences in the methods for processing vacuum bottoms between the base case and MFS case designs result in significant differences in the quantity of net products generated, and hence the overall thermal efficiency of an EDS plant.

There are some differences between the waste streams generated from the EDS base case and MFS case designs, but primarily only in the section that processes the vacuum bottoms. For example, waste streams from the Flexicoking operation in the MFS case are generated at approximately half the rate as corresponding waste streams in the base case design, because only 50% of the vacuum bottoms are fed to Flexicoking in the MFS case. Also, there are waste streams generated from partial oxidation of vacuum bottoms in the MFS case but not in the base case, and waste streams generated from steam reforming of light hydrocarbon gases in the base case but not in the MFS case.

2.1.3 Bottoms Recycle Case

Exxon's bottoms recycle design incorporates reycle of some vacuum bottoms and all vacuum gas oil, as well as hydrogenated donor solvent to the liquefaction reactor. The vacuum gas oil and vacuum bottoms recycle is intended to give these heavy materials longer residence time in the liquefaction reactor in order to convert them to lighter liquids, and reduce the production of heavy materials. The vacuum gas oil is a sidestream recovered from the vacuum distillation column, and the vacuum bottoms are the heavy residues obtained from the same column. This column is designed to recover liquid products from the higher boiling fractions of the coal liquefaction product. Distillation is carried out at low pressures because the boiling temperature decreases with a lowering of pressure. As discussed previously, the bottoms recycle case is covered only briefly in the PCTM.

2.2 EFFECTS OF MAJOR VARIABLES ON DISCHARGES/CONTROLS

An EDS plant is a very complex, integrated facility. There are many variables that affect the emissions from the plant and the approach in controlling emissions. The changes in emissions, of course, could also impact the cost and performance of applicable control technologies. The following provides some brief discussion on some of the major variables.

2.2.1 Effect of Coal Type

A wide range of coals, including bituminous, subbituminous, and lignite, have been liquified using the EDS process. In general, higher rank coal produces more liquid and provides a better service factor (operability) than lower rank coal. Without recycle of bottom products, bituminous coals studied were found to produce 39-46% liquid products, subbituminous coals about 38% liquid products, and the lignite about 36% liquid products (4). These liquid yields are given as weight percent of coal feed on a dry, ash-free basis.

The coal feedstock selected for examination in both the base case and the MFS case designs is an Illinois No. 6 coal, which is a high sulfur bituminous coal. This coal was selected because it is the design coal for the 1975/1976 EDS commercial plant study design (3), and subsequent updates of the EDS commercial plant study design (5-8). Illinois No. 6 coal has also been extensively tested by Exxon Research and Engineering Company (ER&E) in the laboratory, and a considerable amount of process and waste stream characterization data are available. Data on the composition of the Illinois No. 6 coal considered in the PCTM are presented in Table 2-1. This representative composition of an Illinois coal is that which was used by Exxon in their conceptual design, and is that on which the flows/material balances in this PCTM are based. In Table 2-2, data on the trace element contents of the Illinois No. 6 coal are presented.

Using a different type of coal will affect emissions in all three media (air, water and solid). The major factors which can influence the flow rates and characteristics of waste streams, and therefore the selection, costs and performance of control techniques, are discussed as follows.

TABLE 2-1. CHARACTERISTICS OF EDS COAL FEEDSTOCK: REPRESENTATIVE ILLINOIS NO. 6 COAL CONSIDERED IN PCTM

Coal Composition		Dry Wt %	"As Received" Wt %
C		69.9	58.2
H		5.2	4.3
O (By difference)		10.1	8.4
N		1.2	1.0
	Pyritic	1.2	1.0
S	Sulfate	0.1	0.1
	Organic	3.1	2.6
Cl		0.1	0.1
Ash		9.1	7.6
Water		-	16.7
TOTAL		100.0	100.0

	Wt %	
Ash Composition	Dry	SO_3 Free
P_2O_5	1.08	1.10
SiO_2	51.64	52.29
Fe_2O_3	19.37	19.61
Al_2O_3	18.17	18.40
TiO_2	.87	.89
CaO	3.15	3.19
MgO	1.23	1.24
SO_3	1.57	-
K_2O	2.19	2.22
Na_2O	1.07	1.08
TOTAL	100.34	100.02

Data source: Reference (7).

TABLE 2-2. TRACE ELEMENT CONCENTRATIONS OF REPRESENTATIVE ILLINOIS NO. 6 COAL CONSIDERED IN PCTM

Trace Element	Wt % of Ash	Trace Element	Wt % of Ash
Al	9.74	Mg	0.748
As	0.0050	Mn	0.0447
B	0.114	Mo	0.0078
Ba	0.0936	Na	0.801
Be	0.0013	Ni	0.0185
Br	0.0126	P	0.480
Ca	2.28	Pb	0.0228
Cd	0.0034	Rb	0.0135
Ce	0.0110	Sb	0.0008
Cl	1.35	Sc	0.0022
Co	0.0056	Se	0.0019
Cr	0.0169	Si	24.4
Cs	0.0010	Sm	0.0010
Cu	0.0110	Sn	0.0040
Eu	0.0002	Sr	0.0304
F	0.0531	Ta	0.0001
Fe	13.7	Th	0.0019
Ga	0.0026	Ti	0.534
Ge	0.0047	Tl	0.0006
Hf	0.0004	U	0.0013
Hg	0.0002	V	0.0278
K	1.84	Zn	0.354
La	0.0059	Zr	0.0438

Data source: Values for Al, Ca, Fe, K, Mg, Na, P, Si, and Ti are based on ER&E EDS commercial plant design data (7). Values for other trace elements were from Illinois No. 6 coal analysis determined by Gluskoter, et al. (9).

Coal Sulfur. The sulfur in the coal feedstock will eventually end up in acid gas streams, ash or products. The relative distribution of the sulfur in these streams is affected by the reactivity of the coal, product slate, operating conditions of the various units (such as the liquefaction reactor and distillation column) and other design features of the plant (such as process selection for handling vacuum bottoms). There are insufficient data to predict the exact distribution of sulfur in these streams. However, it is reasonable to assume the bulk of the sulfur will appear in the acid gases, i.e., higher sulfur coal will have higher levels of H_2S and other sulfur species in the acid gas streams.

Coal Moisture. As discussed earlier, the coal-solvent slurry feed to the liquefaction reactor is pre-dried to 4% moisture on dried coal basis. Thus, coals with higher moisture content (lower rank coals) will generate larger quantities of wastewater from the slurry drier area.

Coal Ash. Coal ash will eventually end up as ash from Flexicoking operation or bottom partial oxidation (gasification) operation, which is the major solid waste stream from an EDS facility. Thus, higher ash coal will require larger solid waste disposal/management capacity. However, leachabilities of various pollutants (especially trace metals) from Flexicoking or gasifier ash may be highly coal specific and not be easily predictable.

Coal Reactivity. It is known that coal reactivity depends on the solvent, the operating conditions of the liquefaction reactor, and the composition and rank of the coal. Coal reactivity will affect the distribution and composition of liquid and gaseous products, as well as the generation rate and composition of various waste streams. For example, lower reactivity results in the generation of larger quantities of vacuum bottoms (containing higher proportions of unreacted coal), and therefore higher generation rates for waste streams from processing of vacuum bottoms (e.g., by Flexicoking). Also, the differences in coal reactivity could result in the presence of different high molecular weight organics in gaseous waste streams, wastewater streams, solid waste streams such as sludges, and fugitive organic emissions. These high molecular weight organics could be refractory in nature, difficult to destroy in flares and incinerators, and resistant to wastewater treatment. There are, however, no publicly available data to indicate the effects of coal reactivity on waste streams.

2.2.2 Effects of EDS Plant Design

EDS is still an evolving process. Several designs have been considered and some have been tested in the EDS pilot scale unit. The two configurations considered in this manual, the base case and the MFS case are the most advanced in development. The major difference between these two designs is the method of hydrogen production. The effect of this on the waste streams is briefly discussed below. A more detailed discussion, which includes tables summarizing the stream characteristics of the two designs, will be presented in Section 3. In addition to the base and MFS case, Exxon Research and Engineering has recently completed a study on recycling some vacuum bottoms and vacuum gas oil to generate more light liquid products (11). The potential impact of this is also briefly discussed below.

Hydrogen Production

Both the base and MFS cases obtain a portion of the make-up hydrogen requirement by cryogenic separation of gas streams from liquefaction, donor solvent hydrogenation, product distillation, donor solvent fractionation, and Flexicoking areas. However, the balance of the hydrogen requirement is generated by steam reforming for the base case, while the MFS case employs partial oxidation of half of the vacuum-bottoms slurry.

Base case hydrogen generation involves steam reforming of light hydrocarbon gases, shift conversion of carbon monoxide and steam to carbon dioxide and hydrogen, carbon dioxide removal and removal of residual carbon oxides and nitrogen. Feed streams to the reformer are the C_3^- hydrocarbons from cryogenic hydrogen recovery, and the C_1 and C_2 hydrocarbons from light ends recovery. Hydrogen production for the MFS case involves syngas generation in the partial oxidation step, shift conversion of the syngas, acid gas removal and compression. Approximately half of the vacuum-bottoms slurry produced in the liquefaction/distillation units will be processed via partial oxidation into hydrogen while the balance will be processed through Flexicoking to produce additional liquid and gas products.

There are several differences between the waste streams generated from hydrogen production in the base case and MFS cases. First, since only approximately half of the vacuum-bottoms slurry will be processed by Flexicoking in the MFS case, waste streams from Flexicoking in this case will only be generated at approximately half the rate as in the base case. Second, there are three major waste streams in the MFS case that are not present in the base case, due to the addition of the partial oxidation unit. These are ash from partial oxidation process, acid gas from AGR system in the partial oxidation area, and gasifier process condensate from the partial oxidation area. The acid gas stream from partial oxidation is different from acid gas streams generated from other process areas, in that the H_2S/CO_2 ratio of this stream is relatively low and might necessitate the application of different control techniques. This is because in partial oxidation, the CO_2 resulting from the shift conversion is removed simultaneously with the H_2S from partial oxidation. Also, the partial oxidation (gasification) process included in the design operates at a high temperature (∿1670°K) (12). Thus, the process condensate is low in organics compared to most other EDS wastewater streams. The gasifier ash is essentially composed of uncombustible, inert material from coal (12). The combined gasifier ash and Flexicoking ash from the MFS case design is, therefore, similar in properties and generation rate to the Flexicoking ash from the base case design.

Recycle of Vacuum Gas Oil

The EDS Bottom Recycle Study Design (11) recently completed by the Exxon Research and Engineering Company involves recycle of the vacuum gas oil from the vacuum distillation column back to the liquefaction reactor. This mode of operation increases the production of light end products (12), and will likely affect the process wastewaters and gaseous waste streams generated.

Compared to the MFS and base case designs, the Bottom Recycle design requires more severe operating conditions in various units. For example, slurry drier temperature and residence time is increased from 408°K (275°F) and 40 minutes to 422°K (300°F) and 55 minutes, respectively; the hydrogen partial pressure at the reactor inlet is increased from 8.4 MPa (120C psig) to

11.4 MPa (1640 psig); and the liquefaction pressure is increased from 13.9 MPa (2000 psig) to 17.3 MPa (2500 psig). These changes in operating conditions will likely affect the amount of reaction water generated, the quantity and nature of organics and NH_3 in the wastewater stream, and the H_2S distribution in the various gaseous streams. Flexicoking is eliminated as a means for processing vacuum bottoms, thereby altering the identity and composition of streams from bottom processing. There are no data publicly available on the characteristics of any of the waste streams generated in the Bottoms Recycle design.

2.2.3 Effects of Product Slate

The exact product slate from an EDS plant is dependent upon the process design configuration, the reactivity of coal, and the process operation conditions. Table 2-3 presents the net product slate for the base and MFS cases, assuming 27.2 Gg (30,000 tons) per stream day of coal feed on an "as received" basis, and excluding coal being used in steam and power boilers. As discussed before, the base case is designed to produce a maximum amount of C_4^+ liquids; the C_3^- hydrocarbon gas is steam reformed to generate plant hydrogen. In the MFS case, hydrogen is produced via gasification of a portion of the vacuum bottoms; this allows the C_3^- hydrocarbon gas produced within the plant to be sold as an additional product.

As will be apparent from the comparison of the base and MFS streams/compositions in Section 3, the changes in design and operating conditions that effect a change in product slate can cause some variations in the compositions of some streams, and can sometimes result in the appearance of different streams in different designs. However, these differences are generally relatively minor; similar types of streams with similar general compositions will normally be present.

2.3 CAPITAL AND OPERATING COSTS FOR UNCONTROLLED EDS PLANTS

In order to assess the relative impact of the costs of pollution control on overall plant cost, baseline costs for "uncontrolled" plants were developed. Both capital and operating costs for uncontrolled EDS plants were estimated. The total facility cost (total process cost) is the sum of: installed equipment cost (IEC), contractor's overhead and fee (3% of IEC), engineering and construction (25% of IEC), and contingency (20% of IEC). The total depreciable

TABLE 2-3. NET PRODUCT SLATES FOR EDS COMMERCIAL PLANT (STREAM DAY BASIS)

Net Product	Base Case	MFS Case
Pipeline Gas	--	1.731 Gg (11,900 BFOE)
C_3 LPG	0.236 Gg (1,780 BFOE)*	0.334 Gg (2,570 BFOE)
C_4 LPG	0.227 Gg (1,680 BFOE)	0.235 Gg (1,740 BFOE)
Naphtha	2.880 Gg (20,100 BFOE)	2.824 Gg (19,710 BFOE)
Fuel Oil	6.033 Gg (36,680 BFOE)	5.788 Gg (35,160 BFOE)
Total	9.376 Gg (60,240 BFOE)	10.912 Gg (71,080 BFOE)
By-products		
Sulfur	0.885 Gg	0.907 Gg
Ammonia	0.159 Gg	0.089 Gg
Crude Phenols	0.059 Gg (55 m^3)	0.059 Gg (55 m^3)

*1 barrel fuel oil equivalent (BFOE) = 6,050,000 Btu (LHV).

Data source: Reference 13.

Basis: 27.2 Gg (30,000 tons) coal per stream day conceptual commercial design prepared by Exxon.

Product Property Basis

Pipeline Gas	63.0 vol % methane, 22.3 vol % C_2, 2.7 vol % C_3^+ hydrocarbons, 5.4 vol % N_2, 3.4 vol % H_2, and 3.2 vol % CO. Lower heating value of 40.3 MJ/Nm^3 (1,024 Btu/SCF).
C_3 LPG	95.8 wt % C_3, 0.9 wt % C_2^-, 3.3 wt % C_4^+.
C_4 LPG	95.1 wt % C_4, 3.2 wt % C_3^-, 1.7 wt % C_5^+.
Naphtha	Petanes to 350°F normal boiling range; contains 0.43 wt % sulfur and 0.06 wt % nitrogen. Uses: Downstream processing (hydrotreating) required prior to use as gasoline feedstock or in aromatics production.
Fuel Oil (Base Case)	19.0 wt % C_5/400°F, 34.0 wt % 400/700°F, 29.1 wt % 700/1000°F, and 17.9 wt % 1000°F+; contains 0.54 wt % sulfur and 0.77 wt % nitrogen. Uses: Low sulfur fuel oil, refine into stationary turbine fuel, others.
Fuel Oil (MFS Case)	19.2 wt % C_5/400°F, 35.2 wt % 400/700°F, 31.6 wt % 700/1000°F, and 14.0 wt % 1000°F+; contains 0.51 wt % sulfur and 0.75 wt % nitrogen. Uses: As base case fuel oil above.

investment (TDI) is the sum of total facility cost and interest during construction (22.56% of total facility cost). The total capital investment is the sum of total depreciable investment and working capital. Startup costs and land costs are not included. Working capital is assumed to be equivalent to 60 days of inventory of coal and catalysts and chemicals; there is no separate allowance for product inventory and accounts receivable. The operat-cost is the sum of: coal cost, cost of catalysts and chemicals, operating labor cost ($11.00 per hour), operating supervision cost (15% of total operating labor), maintenance cost (2% of capital investment), maintenance supervision cost (5% of maintenance cost), laboratory charges (5% of operating labor), plant general overhead (50% of operating labor), plant G&A overhead (15% of total labor plus maintenance), and contingency (20% of total direct and indirect costs, excluding the cost of coal). The annualized cost is the sum of annual operating cost and capital charges (17.2% of total depreciable investment). The cost methodology used is discussed in greater detail in Section 4.1.3.

2.3.1 Base Case

The installed cost of the uncontrolled plant was estimated by cost data supplied by Exxon. In Table 1-IV-11 of Reference 13, Exxon estimated an installed cost of $1538 million for a plant corresponding to the Illinois coal base case design, with 27.2 Gg (30,000 tons) of "as received" coal feed per stream day and using 4th quarter of 1978 as the cost basis. Included in the plant cost (installed equipment cost) were: $15.9 million for sulfur plant and sulfur plant tail gas cleanup unit, $7.9 million for sour water treating and ammonia recovery, $10.2 million for phenol extraction, $40.1 million for wastewater treatment, $17.0 million for handling of waste solids from the steam boiler (ash and FGD sludge), and $5.2 million for waste solids disposal landfill. The sum of the pollution control cost of $96.3 million was subtracted from the plant cost to obtain an estimated uncontrolled plant cost of $1441.7 million. Acid gas removal and cooling water facilities were considered to be process related functions and included in the uncontrolled plant cost. The plant cost also included all off-site facilities except the coal-fired power plant, as Exxon assumed power for plant use will be purchased. After adjustment for inflation, the uncontrolled plant installed cost was estimated to be $1640 million on 1980 basis.

As shown in Table 2-4, the total capital investment for the uncontrolled EDS base case plant - starting with the $1640 million installed cost, and adding other factors listed above - is $3190 million. Also added to the capital investment cost is the cost of a 268 MW coal-fired power plant, costed at $500/kW. This is in contrast to the original Exxon commercial design, in that the design/costs in the PCTM assume that plant power is generated on site rather than being purchased. The size of the power plant is based on the operating power requirement of 223 MW plus 20% for contingency. Working capital is based on the dollar equivalent of 60 days of coal (1,846 Gg*), and catalysts and chemicals at the plant.

Annualized operating costs for the uncontrolled EDS base case plant, as presented in Table 2-5, were based on a 0.9 plant stream factor. The amount of coal consumed is the sum of coal feed to the liquefaction section (1134 Mg/hr), the power plant (112 Mg/hr), and the steam boiler (35.3 Mg/hr). Annualized cost for coal was computed on the basis of $1.14/GJ ($1.20/MM Btu). Cost for catalysts and chemicals was estimated to be $12 million/year (3rd quarter 1978, 0.85 plant steam factor) by Exxon (13). This cost was adjusted to $13 million by correction for inflation, a higher plant stream factor of 0.9, and allowance for catalysts and chemicals used for pollution control (assumed to be 6%, the same factor used in deriving the uncontrolled plant cost). The operating labor requirement of 580 personnel was estimated using figures provided in the same 1981 Exxon report. Operating labor cost was calculated at $11.00/manhour. The remaining annualized operating costs presented in Table 2-5 were estimated by using factored methods. For example, maintenance cost was assumed to be 2% of capital investment, laboratory charges were assumed to be 5% of operating labor, etc. The factors used are described earlier in this section. The total annualized cost of $960 million is equivalent to $305/m^3 ($48.5/bbl), based on a production rate of 9,580 m^3 (60,240 bbl) fuel oil equivalent of C_3^+ hydrocarbon liquids per stream day.

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* All coal feed rates are given on "as received" basis, unless otherwise stated.

TABLE 2-4. CAPITAL COSTS FOR UNCONTROLLED EDS BASE CASE DESIGN COMMERCIAL PLANT

Item	Cost (Dollars)
Installed Equipment Cost	$1,640.0 \times 10^6$
Contractors Overhead and Fee	49.2
Engineering and Construction	410.0
Contingency	328.0
	$2,427.2 \times 10^6$
Coal-Fired Power Plant	134.0
Subtotal, Facility Cost	$2,561.2 \times 10^6$
Interest During Construction	577.8
Total Depreciable Investment (TDI)	$3,139.0 \times 10^6$
Working Capital	52.8
Total Capital Investment	$3,190 \times 10^6$

Cost basis: 1980 dollars.

TABLE 2-5. ANNUALIZED COSTS FOR UNCONTROLLED EDS BASE CASE DESIGN COMMERCIAL PLANT

Item	Annualized Cost, ($$10^6$)
Coal	276.29
Raw water	0.01
Catalysts and Chemicals	13.00
Operating Labor	13.27
Supervision	1.99
Maintenance Labor and Material	63.79
Maintenance Supervision	3.19
Plant Overhead	6.64
Plant G&A	12.34
Laboratory Charges	0.66
Contingency	22.98
Total, Operating Cost	414.16
Capital Charges (0.172 of TDI)	539.91
Total Annualized Cost	960

2.3.2 MFS Case

The total erected cost of the MFS plant was estimated by Exxon to be 91.84% of the total erected cost of the base case plant (13). Using 4th quarter of 1978 as the cost basis, an installed cost of $1412.5 million was estimated for an MFS plant processing 27.2 Gg (30,000 tons) of "as received" Illinois coal per stream day. In Table 2-I-1 of Reference 13, Exxon estimated the same pollution control cost for the base case and MFS case designs, within 1 to 2 percent. Therefore, the sum of the pollution control cost of $96.3 million was subtracted from the plant cost to obtain an estimated uncontrolled plant cost of $1316.2 million. As in the base case, this plant cost included all off-site facilities except the coal-fired power plant, as Exxon assumed power for plant use will be purchased. After adjustment for inflation, the uncontrolled plant cost was estimated to be $1497.4 million on 1980 basis. As shown in Table 2-6, the capital investment for the uncontrolled EDS MFS plant is $2890 million. The lower capital investment cost for the MFS design is the result of the lower physical plant cost and the smaller power plant (192 MW) requirement. The size of the power plant is based on the operating power requirement of 160 MW plus 20% for contingency.

Annualized operating costs for the uncontrolled EDS MFS plant are presented in Table 2-7. In the MFS design, the amount of coal consumed is 1134 Mg/hr for the liquefaction section, 68.7 Mg/hr for the power plant, and 92.8 Mg/hr for the steam boiler. Cost for catalysts and chemicals was estimated to be $10.8 million/year by Exxon (13). This cost was adjusted accordingly as in the base case. Operating labor requirement for the MFS case design was assumed to be the same as for the base case design (13). The remaining annualized costs presented in Table 2-7 were estimated by using factored methods described earlier. The total annualized cost for the MFS plant is $910 million. This is equivalent to a unit cost of $242/m^3 ($38.5 bbl), based on a production rate of 11,300 m^3 (71,080 bbl) fuel oil equivalent per stream day of C_3^+ hydrocarbon liquids and C_1/C_2 pipeline gas.

TABLE 2-6. CAPITAL COSTS FOR UNCONTROLLED EDS MFS CASE DESIGN COMMERCIAL PLANT

Item	Cost (Dollars)
Installed Equipment Cost	1,497.4 x 10^6
Contractors Overhead and Fee	44.9
Engineering and Construction	374.4
Contingency	299.5
	2,216.2 x 10^6
Coal-Fired Power Plant	96.0
Subtotal, Facility Cost	2,312.2 x 10^6
Interest During Construction	521.6
Total Depreciable Investment (TDI)	2,833.8 x 10^6
Working Capital	53.8
Total Capital Investment	2,890 x 10^6

TABLE 2-7. ANNUALIZED COSTS FOR UNCONTROLLED EDS COMMERCIAL PLANT

Item	Annualized Cost ($$10^6$)
Coal	282.89
Raw Water	0.01
Catalysts and Chemicals	11.63
Operating Labor	13.27
Supervision	1.99
Maintenance Labor and Material	57.71
Maintenance Supervision	2.89
Plant Overhead	6.64
Plant G&A	11.38
Laboratory Charges	0.66
Contingency	21.23
Total, Operating Cost	410.30
Capital Charges (0.172 of TDI)	487.41
Total Annualized Cost	900

3. DETAILED PROCESS DESCRIPTION AND WASTE STREAM CHARACTERIZATION

In this section, the process and auxiliary operations associated with the Exxon Donor Solvent (EDS) process and the characteristics of uncontrolled waste streams resulting from these operations are described. The material presented in Section 3 is used in evaluating pollution control options in Section 4.

Throughout Section 3, the flow rates of waste streams and process streams presented correspond to an EDS commercial plant processing 1,134 Mg/hr (30,000 tons per stream day) of "as received" Illinois No. 6 coal in the liquefaction area. This EDS plant produces about 9,580 m^3 (60,240 barrels) fuel oil equivalent* per stream day of liquid products for the base case design, or 11,300 m^3 (71,080 barrels) fuel oil equivalent per stream day of liquid and gas products for the Market Flexibility Sensitivity (MFS) design. The stream compositions presented throughout Section 3 represent the specific design/operating conditions reflected in the base and MFS designs. These compositions could vary in practice due to uncertainty in the estimates, and due to the variations in coal type and plant design/operating conditions. The commercial plant configurations, and the stream flows and compositions are based largely upon conceptual designs prepared by Exxon.

To facilitate the discussion, a detailed description of the overall EDS process is first presented in Section 3.1. This is followed in Section 3.2 by listings of waste streams and selected process streams for the base case and Market Flexibility Sensitivity (MFS) case designs, the two designs considered in detail in this document. In Section 3.3, detailed descriptions of discharge sources and uncontrolled discharge streams are provided by process area. For discussion purposes, the process operations covered are divided into:

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* 1 barrel fuel oil equivalent (BFOE) = 6,050,000 Btu (LHV)

coal preparation, coal liquefaction, product separation and purification, and processing of liquefaction residue/hydrogen production plant. Auxiliary operations covered are divided into: raw water treatment, steam and power generation, cooling system, oxygen production, and product/by-product storage. Product streams and fugitive emissions are covered in individual subsections. In Section 3.4, characteristics of uncontrolled discharges and product streams are summarized in tabular form by medium and by process area. Finally, cross-references are provided in Section 3.5 for ready reference of applicable control technologies described in Section 4 and the Appendix to waste streams described in Section 3.

As discussed in Section 1, the "uncontrolled discharge streams" discussed in the manual are waste streams prior to the application of any controls. As such, these streams are generally not true discharge streams, since, in a commercial EDS facility, control techniques would normally be applied.

3.1 DETAILED DESCRIPTION OF OVERALL EDS PROCESS

The Exxon Donor Solvent (EDS) is a noncatalytic process that liquefies coal by reacting the coal with hydrogen in a medium of a hydrogen donor solvent obtained from coal-derived distillates. The donor solvent assists in the transfer of hydrogen to the coal, maximizes the yield of desired liquid products and prevents the conversion of the reaction coal into char and asphalt-like materials. Two process configurations, based on Exxon's Study Design Update completed in 1981, are described in this manual. These two process configurations are referred to as the base case design and the Market Flexibility Sensitivity (MFS) design (5,7,13). The method for hydrogen production is the major difference between these two designs. In the base case, hydrogen for the liquefaction reaction is produced by steam reforming of light hydrocarbon gases generated in the plant. In the MFS design, the hydrogen is produced by partial oxidation of about 50 percent of the vacuum bottoms, the heavy residual material remaining after vacuum distillation of liquefaction products has removed all of the desired liquids. Hence, the light hydrocarbon gases made available by elimination of steam reforming can be sold as pipeline gas.

Because of the differences in design, the overall average energy efficiency* is 56% for the base case compared with 63% for the MFS case, when Illinois No. 6 is used as the feed coal (13).

More recently, Exxon has completed the preliminary conceptual design of an EDS plant feeding Wyoming subbituminous coal with recycle of all vacuum gas oil (VGO) and of some vacuum bottoms (11). In this design, the VGO - the heaviest liquid produced from the plant, above vacuum bottoms - is recycled to the liquefaction reactor in order to convert it to lighter products. This Bottoms Recycle Study design is not addressed in detail in the PCTM because: 1) the level of design detail for this design is less than that of the base case and MFS case designs; and 2) waste stream characterization and pollution control evaluation studies for the EDS process were completed before information on this particular design became available. However, the primary differences between this design and the base case and MFS case designs are addressed briefly in Section 3.1.3.

3.1.1 Base Case

The base case configuration for the EDS process, designed to maximize the production of C_4^+ liquids, is shown schematically in Figure 3-1. The major process areas include: coal preparation, coal liquefaction, product separation and purification, and liquefaction residue processing/hydrogen production. Auxiliary operations shown include raw water treatment, steam generation, power generation, cooling system, and product/by-product storage. Pollution control operations such as sulfur recovery, ammonia recovery, phenol extraction, and wastewater treatment are also included in the block flow diagram to indicate the flows of waste streams into various treatment areas.

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* Energy efficiency $= \frac{A}{B + C}$

where A = heating value of net products + by-products
B = heating value of process coal + off-site coal
C = heating value of fossil fuel to generate required electric power

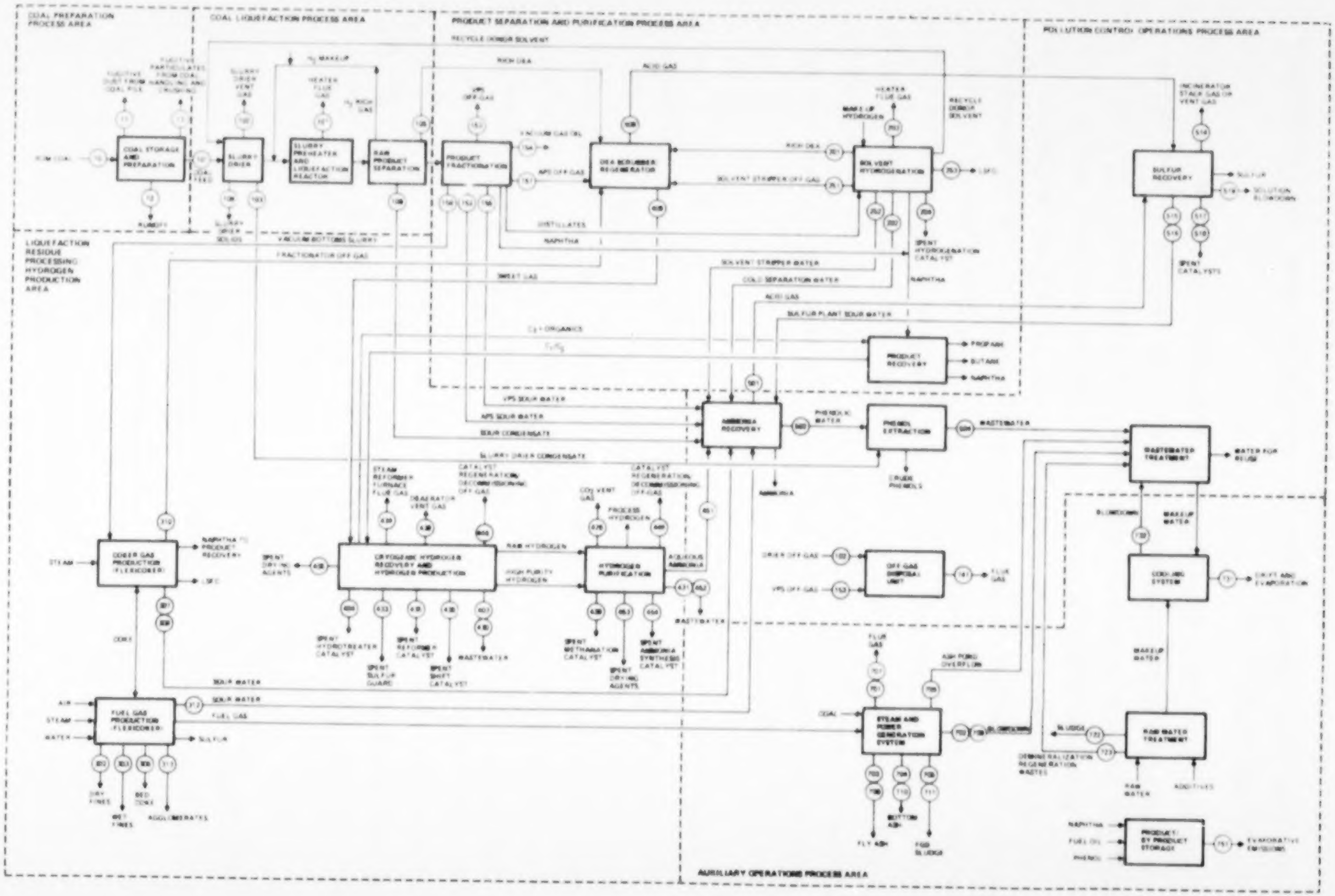
COAL PREPARATION PROCESS AREA
COAL LIQUEFACTION PROCESS AREA
PRODUCT SEPARATION AND PURIFICATION PROCESS AREA
POLLUTION CONTROL OPERATIONS PROCESS AREA
LIQUEFACTION RESIDUE PROCESSING HYDROGEN PRODUCTION AREA
AUXILIARY OPERATIONS PROCESS AREA
ROM COAL
COAL STORAGE AND PREPARATION
FUGITIVE DUST FROM COAL PILE
FUGITIVE PARTICULATES FROM COAL HANDLING AND CRUSHING
RUNOFF
COAL FEED
SLURRY DRIER
SLURRY DRIER VENT GAS
SLURRY DRIER SOLIDS
H_2 MAKEUP
SLURRY PREHEATER AND LIQUEFACTION REACTOR
HEATER FLUE GAS
H_2 RICH GAS
RAW PRODUCT SEPARATION
RECYCLE DONOR SOLVENT
RICH DEA
PRODUCT FRACTIONATION
VPS OFF GAS
VACUUM GAS OIL
APS OFF GAS
DEA SCRUBBER REGENERATOR
ACID GAS
RICH DEA
SOLVENT STRIPPER OFF GAS
SOLVENT HYDROGENATION
MAKE UP HYDROGEN
HEATER FLUE GAS
RECYCLE DONOR SOLVENT
LSFO
SPENT HYDROGENATION CATALYST
SULFUR RECOVERY
INCINERATOR STACK GAS OR VENT GAS
SULFUR
SOLUTION BLOWDOWN
SPENT CATALYSTS
VACUUM BOTTOMS SLURRY
FRACTIONATOR OFF GAS
DISTILLATES
NAPHTHA
SWEET GAS
SOLVENT STRIPPER WATER
COLD SEPARATION WATER
ACID GAS
SULFUR PLANT SOUR WATER
NAPHTHA
C_3 + ORGANICS
C_1/C_2
PRODUCT RECOVERY
PROPANE
BUTANE
NAPHTHA
VPS SOUR WATER
APS SOUR WATER
SOUR CONDENSATE
SLURRY DRIER CONDENSATE
AMMONIA RECOVERY
AMMONIA
PHENOLIC WATER
PHENOL EXTRACTION
CRUDE PHENOLS
WASTEWATER
WASTEWATER TREATMENT
WATER FOR REUSE
STEAM
COKER GAS PRODUCTION (FLEXICOKER)
NAPHTHA TO PRODUCT RECOVERY
LSFO
COKE
AIR
STEAM
WATER
FUEL GAS PRODUCTION (FLEXICOKER)
SULFUR
DRY FINES
WET FINES
BED COKE
AGGLOMERATES
SOUR WATER
SOUR WATER
FUEL GAS
SPENT DRYING AGENTS
CRYOGENIC HYDROGEN RECOVERY AND HYDROGEN PRODUCTION
STEAM REFORMER FURNACE FLUE GAS
DEAERATOR VENT GAS
CATALYST REGENERATION/DECOMMISSIONING OFF GAS
SPENT HYDROTREATER CATALYST
SPENT SULFUR GUARD
SPENT REFORMER CATALYST
SPENT SHIFT CATALYST
WASTEWATER
RAW HYDROGEN
HIGH PURITY HYDROGEN
HYDROGEN PURIFICATION
CO_2 VENT GAS
PROCESS HYDROGEN
CATALYST REGENERATION/DECOMMISSIONING OFF GAS
AQUEOUS AMMONIA
WASTEWATER
SPENT METHANATION CATALYST
SPENT DRYING AGENTS
SPENT AMMONIA SYNTHESIS CATALYST
DRIER OFF GAS
VPS OFF GAS
OFF GAS DISPOSAL UNIT
FLUE GAS
BLOWDOWN
MAKEUP WATER
COOLING SYSTEM
DRIFT AND EVAPORATION
MAKEUP WATER
COAL
STEAM AND POWER GENERATION SYSTEM
FLUE GAS
ASH POND OVERFLOW
BLOWDOWN
FLY ASH
BOTTOM ASH
FGD SLUDGE
SLUDGE
DEMINERALIZATION REGENERATION WASTES
RAW WATER TREATMENT
RAW WATER
ADDITIVES
NAPHTHA
FUEL OIL
PHENOL
PRODUCT/BY PRODUCT STORAGE
EVAPORATIVE EMISSIONS

Figure 3-1. Block flow diagram for EDS commercial plant - base case

In the coal preparation area, cleaned coal from the mines is transferred for storage in stockpiles. The cleaned coal is then reduced from 90% minus 25 mm (1 inch) to minus 8 mesh (2.4 mm) in impact-mill crushers, and fed via enclosed belt conveyors to the slurry drier.

The coal liquefaction area consists of the slurry drier, the liquefaction reactor, and raw product separation operations. In the slurry drier, the crushed coal is mixed with recycle donor solvent. The slurry is then heated by a preheat furnace, and preheated hydrogen is added. The slurry and the hydrogen are then fed to a liquefaction reactor operated at 700-750°K (800-900°F) and 13.9 MPa (2000 psia). In the liquefaction reactor, the polymeric coal molecules are broken down into smaller, lighter molecules through the solvent-aided addition of hydrogen. Most of the organically-bound sulfur is converted to H_2S, which then appears in various process streams. Products from the liquefaction reactor are separated in raw product separation into a gas stream and a liquid/solids stream. The gas stream is further cooled to condense vaporized hydrocarbon liquids, and scrubbed with diethanolamine (DEA) for hydrogen sulfide and carbon dioxide removal. This treated hydrogen-rich gas stream is then compressed for recycle to the liquefaction reactor with make-up hydrogen.

In the product separation and purification area, the liquid/solids stream and the hydrocarbon condensate recovered from the gas stream in raw product separation are fed first to atmospheric and then vacuum fractionators to separate the liquefaction products. Recovered naphtha is sent to light ends processing (product recovery) for stabilization. Vacuum gas oil product is removed as the bottom sidestream of the vacuum fractionator and sent to product storage. The distillates heavier than naphtha but lighter than VGO - consisting of atmospheric and vacuum fractionator sidestreams - are sent to solvent hydrogenation in order to recover the spent solvent and to recover additional liquid products. Hydrogen is catalytically added to the spent solvent before it is recycled to the slurry drier in the liquefaction area, in order to restore its hydrogen - donating ability; this solvent hydrogenation step is a distinguishing feature of the EDS technology. In solvent hydrogenation, the spent solvent distillate is hydrotreated in a fixed-bed catalytic reactor operating at about 12.3 MPa (1785 psia). Effluent from the solvent hydrogenation reactor is cooled and separated into a hydrogen-rich gas and a hydrotreated liquid stream. The gas is scrubbed with DEA, and along with make-up hydrogen, is recycled to the solvent hydrogenation unit. The hydrotreated liquid is fed to the solvent fractionator to recover the hydrogen donor solvent, a hydrotreated low-sulfur fuel oil (LSFO) product, and a naphtha stream.

In the liquefaction residue processing/hydrogen plant area, the heavy bottoms from the vacuum distillation are fed to a Flexicoking* unit with air and steam to produce additional distillate liquid products, a low-Btu fuel gas for process furnaces, and light gases for hydrogen production. Inorganically-bound sulfur largely remains in the vacuum bottoms throughout the liquefaction and product separation steps, and is released as H_2S when the vacuum bottoms are processed in the Flexicoker. The light gases from the Flexicoking unit, in combination with the atmospheric and solvent fractionator off-gases, are scrubbed with DEA and sent to cryogenic hydrogen recovery. The cryogenic hydrogen recovery section separates the high pressure hydrogen purge gases from liquefaction and solvent hydrogenation and the low pressure sweet gas from the DEA absorber for recovery of hydrogen, C_1/C_2, and C_3^+. The C_1/C_2 hydrocarbon gases from cryogenic hydrogen recovery, along with C_1/C_2 hydrocarbon gases from the product recovery section, are then fed to a steam reformer. This is followed by shift conversion and carbon dioxide removal to produce process hydrogen. The hydrogen product is blended with the hydrogen from the cryogenic recovery section, treated for removal of trace carbon oxides and nitrogen, and compressed for use as make-up hydrogen.

3.1.2 Market Flexibility Sensitivity Case

The MFS design for the EDS process is shown schematically in Figure 3-2. As in the base case design, the MFS case also encompasses four major process areas: coal preparation, coal liquefaction, product separation and purification, and liquefaction residue processing/hydrogen production.

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* "Flexicoking" is the service mark for a proprietary process developed by Exxon.

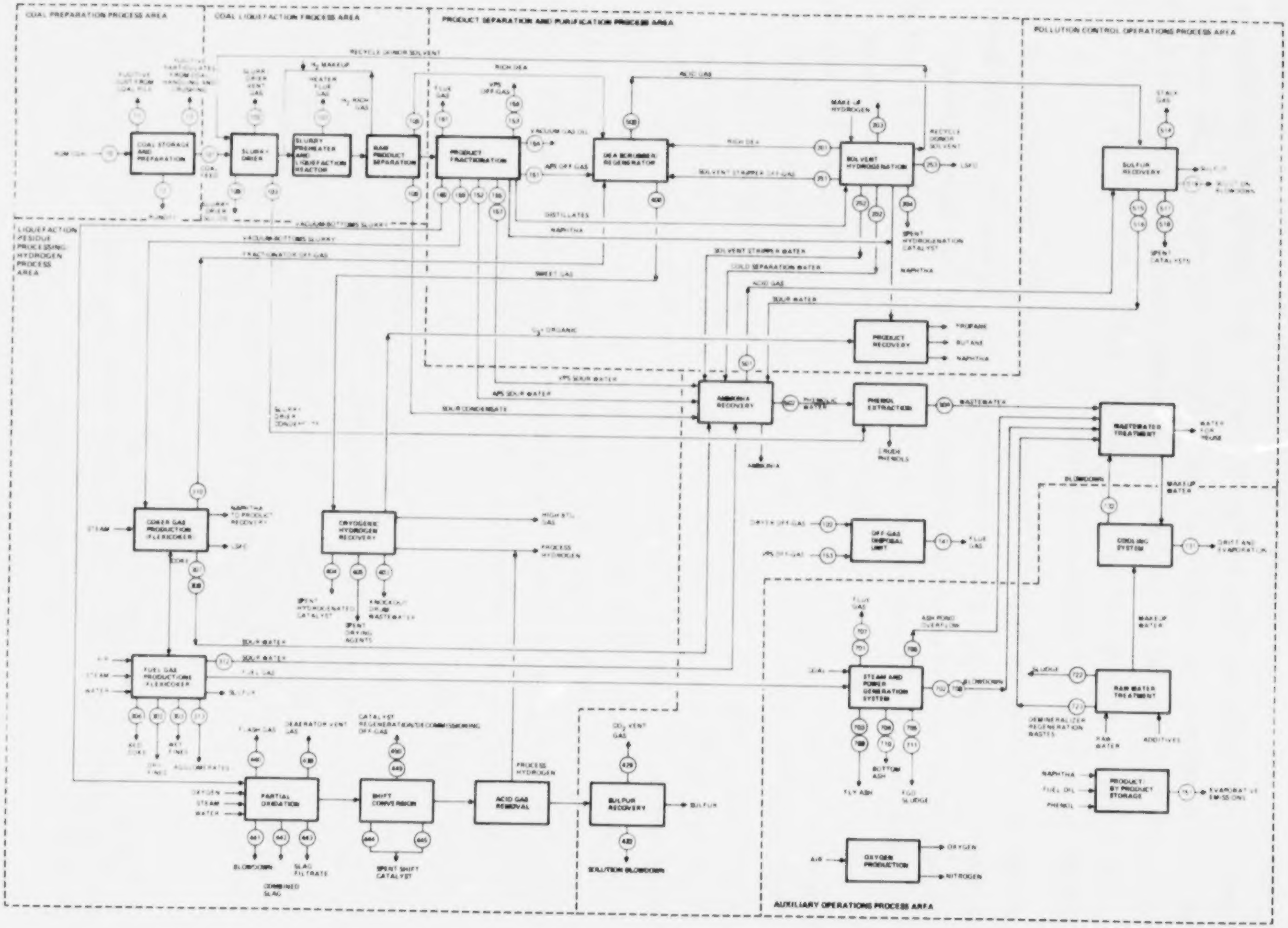
COAL PREPARATION PROCESS AREA
COAL LIQUEFACTION PROCESS AREA
PRODUCT SEPARATION AND PURIFICATION PROCESS AREA
POLLUTION CONTROL OPERATIONS PROCESS AREA
LIQUEFACTION RESIDUE PROCESSING HYDROGEN PROCESS AREA
AUXILIARY OPERATIONS PROCESS AREA
ROM COAL
COAL STORAGE AND PREPARATION
FUGITIVE DUST FROM COAL PILE
FUGITIVE PARTICULATES FROM COAL HANDLING AND CRUSHING
RUNOFF
COAL FEED
SLURRY DRIER
SLURRY DRIER VENT GAS
SLURRY DRIER SOLIDS
SLURRY PREHEATER AND LIQUEFACTION REACTOR
H_2 MAKEUP
HEATER FLUE GAS
H_2 RICH GAS
RAW PRODUCT SEPARATION
RECYCLE DONOR SOLVENT
RICH DEA
PRODUCT FRACTIONATION
FLUE GAS
VPS OFF-GAS
VACUUM GAS OIL
APS OFF GAS
DISTILLATES
NAPHTHA
DEA SCRUBBER/ REGENERATOR
ACID GAS
SOLVENT STRIPPER OFF GAS
MAKE-UP HYDROGEN
SOLVENT HYDROGENATION
RECYCLE DONOR SOLVENT
LSFO
SPENT HYDROGENATION CATALYST
VACUUM BOTTOMS SLURRY
VACUUM BOTTOMS SLURRY
FRACTIONATOR OFF-GAS
SWEET GAS
SOLVENT STRIPPER WATER
COLD SEPARATION WATER
ACID GAS
SOUR WATER
C_3+ ORGANIC
PRODUCT RECOVERY
PROPANE
BUTANE
NAPHTHA
VPS SOUR WATER
APS SOUR WATER
SOUR CONDENSATE
SLURRY DRIER CONDENSATE
AMMONIA RECOVERY
AMMONIA
PHENOLIC WATER
PHENOL EXTRACTION
CRUDE PHENOLS
WASTEWATER
SULFUR RECOVERY
STACK GAS
SULFUR
SOLUTION BLOWDOWN
SPENT CATALYSTS
WASTEWATER TREATMENT
WATER FOR REUSE
BLOWDOWN
MAKEUP WATER
COOLING SYSTEM
DRIFT AND EVAPORATION
STEAM
COKER GAS PRODUCTION (FLEXICOKER)
NAPHTHA TO PRODUCT RECOVERY
LSFO
COKE
CRYOGENIC HYDROGEN RECOVERY
HIGH BTU GAS
PROCESS HYDROGEN
SPENT HYDROGENATED CATALYST
SPENT DRYING AGENTS
KNOCKOUT DRUM WASTEWATER
DRYER OFF-GAS
VPS OFF-GAS
OFF-GAS DISPOSAL UNIT
FLUE GAS
AIR
STEAM
WATER
FUEL GAS PRODUCTION (FLEXICOKER)
SOUR WATER
SOUR WATER
FUEL GAS
SULFUR
BED COKE
DRY FINES
WET FINES
AGGLOMERATES
FLASH GAS
DEAERATOR VENT GAS
CATALYST REGENERATION/DECOMMISSIONING OFF-GAS
OXYGEN
STEAM
WATER
PARTIAL OXIDATION
BLOWDOWN
COMBINED SLAG
SLAG FILTRATE
SHIFT CONVERSION
SPENT SHIFT CATALYST
ACID GAS REMOVAL
PROCESS HYDROGEN
SULFUR RECOVERY
CO_2 VENT GAS
SULFUR
SOLUTION BLOWDOWN
COAL
STEAM AND POWER GENERATION SYSTEM
FLUE GAS
ASH POND OVERFLOW
BLOWDOWN
FLY ASH
BOTTOM ASH
FGD SLUDGE
SLUDGE
RAW WATER TREATMENT
DEMINERALIZER REGENERATION WASTES
RAW WATER
ADDITIVES
MAKEUP WATER
NAPHTHA
FUEL OIL
PHENOL
PRODUCT/ BY PRODUCT STORAGE
EVAPORATIVE EMISSIONS
AIR
OXYGEN PRODUCTION
OXYGEN
NITROGEN

Figure 3-2. Block flow diagram for EDS commercial plant - MFS case

There is no difference between the base case and MFS case designs in the coal preparation and coal liquefaction areas. In product separation and purification, the atmospheric fractionator is followed by two parallel vacuum fractionators of different designs in the MFS case. One vacuum fractionator is designed with a bottoms cut point* of 768°K (920°F) for feed to the Flexicoking unit as in the base case. However, the second vacuum fractionator is designed with a deeper bottoms cut point of 798°K (975°F) for feed to the partial oxidation unit for hydrogen production.

The principal difference between the base case and MFS case designs is the method for hydrogen production. In the MFS design, approximately 50 percent of the vacuum bottoms (i.e., the bottoms product in the second vacuum fractionator described above) are fed to the partial oxidation unit with steam and oxygen. The raw synthesis gas produced is converted to hydrogen through shift conversion and acid gas removal. The remaining 50 percent of the vacuum bottoms (from the first vacuum fractionator) are fed to a Flexicoking unit to produce distillate liquids, a low-Btu fuel gas, and light hydrocarbon gases. Because steam reforming for hydrogen production is no longer required, the light hydrocarbon gases from the Flexicoking unit and from the product separation area are recovered in the cryogenic hydrogen recovery section as a C_1/C_2 high-Btu gas product. Hydrogen from the cryogenic hydrogen recovery section is blended with the treated hydrogen from the partial oxidation section, compressed, and used as make-up hydrogen for the liquefaction reaction.

3.1.3 Bottoms Recycle Case

The Bottoms Recycle case design (with Wyoming subbituminous coal) differs from the base case and MFS case designs primarily in the following features:

- Recycle of some vacuum bottoms and all vacuum gas oil (VGO) as well as donor solvent to the liquefaction reactor. The VGO and vacuum bottoms recycle is intended to give these heavy materials longer residence time in the liquefaction reactor, in order to convert them to lighter liquids, and reduce production of heavy materials. Only the donor solvent is recycled to the liquefaction reactor in the Illinois coal base case and MFS case designs.

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* Bottoms cut point - a temperature limit for the bottoms product; products with boiling points above this temperature limit are obtained in the distillate bottoms.

- Gasification of most of the remaining vacuum bottoms to produce plant hydrogen.
- Firing of the last of the remaining solid vacuum bottoms in conjunction with additional coal, in a hybrid boiler to provide the necessary process heat and plant steam. C_2^- high-Btu gas is used to meet the balance of fuel needs in the plant. Flexicoking units are eliminated.
- More severe operating requirements including: finer coal feed (-20 mesh); higher temperature (422°K) and longer residence time (55 min) in the slurry drier; and higher pressure (17.3 MPa or 2500 psig) longer residence time (60 min), and higher hydrogen partial pressure (11.4 MPa or 1640 psig) in the liquefaction reactor.

As a result of thes design changes, the Bottoms Recycle Study design produces no 850°F+ product, reduces the amount of C_2^- for sale, and delivers a thermal efficiency (with Wyoming coal) of 59%.

3.2 LIST OF UNCONTROLLED AND SECONDARY DISCHARGE STREAMS

A listing of the uncontrolled waste streams by process area and by medium is provided in this section. Secondary discharge streams from pollution control operations are listed as a distinct subset. In addition, to facilitate the tracking of material flows between processing operations, selected process streams are also listed. Uncontrolled waste streams, as the name implies, are waste streams generated by process operations with no pollution controls in place. Secondary discharge streams, on the other hand, are waste streams generated as a result of pollution control. For example, the acid gas generated from sour water stripping/ammonia recovery is a secondary discharge stream.

3.2.1 Base Case

A listing of the waste streams and selected process streams for the Illinois coal base case design is presented in Table 3-1. Uncontrolled discharge streams are listed by process area, and also by medium within each process area. Selected process streams are listed by process area but not by medium. Secondary discharge streams are listed under pollution control operations area because uncontrolled discharge streams are often combined for treatment in a separate plant area.

TABLE 3-1. LISTING OF WASTE STREAMS AND SELECTED PROCESS STREAMS FOR THE EDS ILLINOIS COAL BASE CASE DESIGN

Coal Preparation Process Area	
Uncontrolled Discharge Streams	
Gaseous	
011	Fugitive dust from coal pile
013	Fugitive particulate from coal handling and crushing
Liquid	
012	Coal pile runoff
Process Stream	
010	Run-of-mine coal
Coal Liquefaction Process Area	
Uncontrolled Discharge Streams	
Gaseous	
102	Slurry drier vent gas
107	Flue gas from liquefaction slurry preheat furnace
803	Transient gas from liquefaction reactor
Liquid	
103	Slurry drier cold separator wastewater
106	Liquefaction cold separator wastewater
Solid	
108	Solids accumulated in the slurry drier
Process Streams	
101	Coal feed to slurry dryer
104	Liquefaction cold separator sour gas
Product Separation and Purification Process Area	
Uncontrolled Discharge Streams	

(Continued)

TABLE 3-1. (Continued)

Gaseous	
153	Vacuum fractionator off-gas*
203	Flue gas from solvent hydrogenation fuel preheat furnaces
508	Acid gas from DEA regenerator
Liquid	
152	Atmospheric fractionator overhead drum wastewater
155	Vacuum fractionator wastewater
202	Solvent hydrogenation cold separator wastewater
252	Solvent hydrogenation fractionator overhead drum wastewater
Solid	
204	Spent solvent hydrogenation catalyst
Process Streams	
151	Atmospheric fractionator off-gas
159	Vacuum bottoms slurry to Flexicoking
200	Solvent hydrogenation cold separator vapor
251	Solvent hydrogenation fractionator sour gas
Liquefaction Residue Processing/Hydrogen Production Process Area	
Uncontrolled Discharge Streams	
Gaseous	
304	Flexicoking gasifier/heater sour gas*
426	Vent gas from CO_2 removal
434	Flue gas from hydrogen plant reformer furnaces
438	Hydrogen Plant deaerator vent
446	Regeneration/decommissioning off-gas from reformer catalyst
448	Decommissioning off-gas from methanation catalyst
801	Transient gas from Flexicoking

(Continued)

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* Will normally be desulfurized and burned within the plant as fuel gas.

TABLE 3-1. (Continued)

Liquid	
307	Flexicoking recontacting drum wastewater
308	Flexicoking fractionator overhead drum wastewater
312	Flexicoking heater overhead drum wastewater
403	Knockout drum wastewater in H_2 cryo recovery
430	Blowdown and K.O. drum wastewater from hydrogen generation
431	Catacarb overhead receiver wastewater in hydrogen generation
451	Aqueous ammonia from ammonia synthesis
452	Knockout drum wastewater in ammonia synthesis
Solid	
302	Flexicoking gasifier/heater dry fines
303	Flexicoking gasifier/heater wet fines
306	Flexicoking gasifier/heater bed coke
313	Flexicoking heater/reactor chunks/agglomerates
404	Spent hydrotreater catalyst in H_2 cryo recovery
405	Spent drying agents in H_2 cryo recovery
433	Spent sulfur guard in hydrogen generation
435	Spent reformer catalyst in hydrogen generation
436	Spent shift catalyst in hydrogen generation
439	Spent methanation catalyst in hydrogen generation
453	Spent drying agents in ammonia synthesis
454	Spent ammonia synthesis catalvst
Process Streams	
310	Flexicoking fractionator off-gas
406	Sweet gas from DEA scrubbing
Auxiliary Operations Process Area	
Uncontrolled Discharge Streams	
Gaseous	
701a	Flue gas from steam generation system
707a	Flue gas from power generation system
731	Drift and evaporation from cooling tower
751	Evaporative emissions from product and by-product storage

(Continued)

TABLE 3-1. (Continued)

Liquid	
702	Blowdown from steam generation system
708	Blowdown from power generation system
723	Regeneration wastes from water demineralization
732	Cooling tower blowdown
901	Low quality rain runoff
Solid	
704	Bottom ash from steam generation system
710	Bottom ash from power generation system
722	Raw water treatment sludge
Pollution Control Operations Area	
Secondary Discharge Streams	
Gaseous	
501	Acid gas from sour water stripping/ammonia recovery
510	Tail gas from sulfur plant
512	Off-gas from sulfur plant tail gas treatment unit (prior to incineration)
514	Incinerator stack gas from sulfur plant tail gas treatment unit
525	Flue gas from wastewater sludge incineration
527	Brine concentration off-gas
529	Activated carbon regeneration off-gases
540	Oxidizer vent gas from Stretford process
541	Oxidizer vent gas from Stretford unit in Beavon Sulfur Removal Process (BSRP)
542	Incinerated vent gas from CO_2 removal by Catacarb process
543	Incinerated hydrogen plant deaerator vent
544	Incinerated regeneration/decommissioning off-gas from reformer catalyst
545	Incinerated decommissioning off-gas from methanation catalyst
701	Flue gas from steam generation system (discharge to the atmosphere following controls)
707	Flue gas from power generation system (discharge to the atmosphere following controls)
741	Flue gas from combustion of slurry drier vent gas and vacuum flash off-gas

(Continued)

TABLE 3-1. (Continued)

811	Controlled transient waste gas from Flexicoking
813	Controlled transient waste gas from liquefaction reactor
Liquid	
502	Wastewater from ammonia recovery
504	Wastewater from phenolic extraction
515	Sour water from sulfur recovery plant
516	Sour water from sulfur plant tail gas treatment unit
519	Solution blowdown from sulfur plant tail gas treatment unit
521	Effluent from wastewater treatment
522	Reclaimed water from reuse
706	Ash pond overflow from steam and power generation systems
Solid	
014	Collected coal fines from particulate control in coal preparation
517	Spent sulfur plant catalyst
518	Spent catalyst from sulfur plant tail gas treatment unit
523	Biological sludge from wastewater treatment
524	Sludge from chemical treatment of wastewater
526	Residue from wastewater incineration
528	Residue from brine concentration
530	Recovered sulfur from sulfur plant and tail gas treatment unit
703	Fly ash from steam generation system
705	FGD sludge from steam generation system
709	Fly ash from power generation system
711	FGD sludge from power generation system

3.2.2 MFS Case

A listing of the waste streams and selected process streams for the Illinois coal MFS case design is presented in Table 3-2. There is no difference in the listing of waste streams between the MFS case and base case designs in the coal preparation, coal liquefaction and auxiliary operations process areas. Listing of waste streams from these three process areas is therefore not repeated. In the product separation and purification process area, the MFS case design generates three additional waste streams because a separate vacuum fractionator is provided for the partial oxidation feed. In the liquefaction residue processing/hydrogen plant process area, the waste streams from the MFS case and base case designs are significantly different because of the method of hydrogen production. In the pollution control operations area, the MFS case design generates two additional waste streams because a separate H_2S removal/sulfur recovery unit was assumed for the treatment of acid gas streams containing less concentrated levels of H_2S from the partial oxidation unit.

TABLE 3-2. LISTING OF WASTE STREAMS AND SELECTED PROCESS STREAMS FOR THE EDS ILLINOIS COAL MFS CASE DESIGN*

Product Separation and Purification Process Area	
Uncontrolled Discharge Streams	
Gaseous	
153	Vacuum fractionator off-gas
156	Partial oxidation feed vacuum fractionator off-gas
161	Flue gas from partial oxidation feed vacuum fractionator preheat furnaces
203	Flue gas from solvent hydrogenation fuel preheat furnaces
508	Acid gas from DEA regenerator
Liquid	
152	Atmospheric fractionator overhead drum wastewater
155	Vacuum fractionator wastewater
157	Partial oxidation feed vacuum fractionator wastewater
202	Solvent hydrogenation cold separator wastewater
252	Solvent hydrogenation fractionator overhead drum wastewater
Solid	
204	Spent solvent hydrogenation catalyst
Process Streams	
151	Atmospheric fractionator off-gas
159	Vacuum bottoms slurry to Flexicoking
200	Solvent hydrogenation cold separator vapor
251	Solvent hydrogenation sour gas
Liquefaction Residue Processing/Hydrogen Production Process Area	

(Continued)

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* Listing of waste streams in the coal preparation, coal liquefaction and auxiliary operations is the same as for the base case design and not repeated here.

TABLE 3-2. (Continued)

Uncontrolled Discharge Streams

Gaseous	
304	Flexicoking gasifier/heater sour gas
428	Acid gas from acid gas removal unit in hydrogen purification
438	Hydrogen plant deaerator vent
440	Flash gas from partial oxidation unit
449	Regeneration/decommissioning off-gas from high temperature shift conversion catalyst
450	Regeneration/decommissioning off-gas from low temperature shift conversion catalyst
801	Transient waste gas from Flexicoking
802	Transient waste gas from partial oxidation unit
Liquid	
307	Flexicoking recontacting drum wastewater
308	Flexicoking fractionator overhead drum wastewater
312	Flexicoking gasifier/heater overhead drum wastewater
403	Knockout drum wastewater in H_2 cryo recovery
430	Blowdown and K.O. drum wastewater from hydrogen generation
441	Sour water from partial oxidation unit
443	Slag filtrate from partial oxidation unit
Solid	
302	Flexicoking gasifier/heater dry fines
303	Flexicoking gasifier/heater wet fines
306	Flexicoking gasifier/heater bed coke
313	Flexicoking heater/reactor chunks/agglomerates
404	Spent hydrotreater catalyst in H_2 cryo recovery
405	Spent drying agents in H_2 cryo recovery
442	Slag from partial oxidation unit
444	Spent high temperature shift catalyst in hydrogen generation
445	Spent low temperature shift catalyst in hydrogen generation

(Continued)

TABLE 3-2. (Continued)

Process Streams	
310	Flexicoking fractionator off-gas
406	Sweet gas from DEA scrubbing
Pollution Control Operations Area	
Secondary Discharge Streams	
Gaseous	
429	CO_2 rich vent gas from H_2S removal/sulfur recovery unit in hydrogenation
501	Acid gas from sour water stripping/ammonia recovery
510	Tail gas from sulfur plant
512	Off-gas from sulfur plant tail gas treatment unit (prior to incineration)
514	Incinerator stack gas from sulfur plant tail gas treatment unit
525	Flue gas from wastewater sludge incineration
527	Brine concentration off-gas
529	Activated carbon regeneration off-gases
540	Oxidizer vent gas from Stretford process
541	Oxidizer vent gas from Stretford unit in Beavon Sulfur Removal Process (BSRP)
543	Incinerated hydrogen plant deaerator vent
546	Controlled regeneration/decommissioning off-gas from high and low temperature shift catalysts
701	Flue gas from steam generation system (discharge to the atmosphere following controls)
707	Flue gas from power generation system (discharge to the atmosphere following controls)
741	Flue gas from combustion of slurry drier vent gas and vacuum flash off-gas
811	Controlled transient waste gas from Flexicoking
812	Controlled transient waste gas from partial oxidation unit
813	Controlled transient waste gas from liquefaction reactor

(Continued)

TABLE 3-2. (Continued)

Liquid	
432	Solution blowdown from H_2S removal/sulfur recovery unit
502	Wastewater from ammonia recovery
504	Wastewater from phenolic extraction
515	Sour water from sulfur recovery plant
516	Sour water from sulfur plant tail gas treatment unit
519	Solution blowdown from sulfur plant tail gas treatment unit
521	Effluent from wastewater treatment
522	Reclaimed water from reuse
706	Ash pond overflow from steam and power generation systems
Solid	
014	Collected coal fines from particulate control
517	Spent sulfur plant catalyst
518	Spent catalyst from sulfur plant tail gas treatment unit
523	Biological sludge from wastewater treatment
524	Sludge from chemical treatment of wastewater
526	Residue from wastewater incineration
528	Residue from brine concentration
530	Recovered sulfur from sulfur plant and tail gas treatment unit
703	Fly ash from steam generation system
705	FGD sludge from steam generation system
709	Fly ash from power generation system
711	FGD sludge from power generation system

3.3 DETAILED DESCRIPTION OF DISCHARGE SOURCES BY PROCESS AREA

In this subsection, the sources and characteristics of discharge streams from the EDS process are described in detail. The descriptions are provided according to process area. Fugitive emissions are covered as a separate section because in most cases they are not generated from any one process area alone. Products from the EDS process are also covered in a separate section.

3.3.1 Coal Preparation Area (Area 1)

The amount of coal handled by the coal preparation plant was determined by assessing the coal requirements for the liquefaction reactor, the power boiler, and the steam boiler. These coal requirements are listed in Table 3-3 on an "as received" basis. Part of the plant fuel requirements to produce process steam will be met by plant products and by-products.

The coal preparation area includes coal storage and coal handling operations. A simplified flow diagram indicating all the major process operations and the emission sources is presented in Figure 3-3. As shown in this figure, the only difference between the base case and MFS case designs is the amount of coal handled for steam and power boiler requirements.

TABLE 3-3. COAL FEED REQUIREMENTS - EDS PLANT

Unit	Base Case, Mg/hr	Market Flexibility Sensitivity (MFS) Mg/hr
Liquefaction Reactor	1,134	1,134
Power Boiler	95.8	68.7
Steam Boiler	35.3	92.8

Data Source: Reference 13

3.3.1.1 Coal Storage (Coal Preparation - Area 1)

In the Exxon designs, cleaned and washed Illinois No. 6 coal is received five days per week 24 hours per day from three adjacent mines by three belt conveyors. The incoming coal is stored in two parallel live storage piles. The two piles together have a combined capacity of 5 days throughput of coal. A 30-day dead storage is built up and maintained. Two 100% capacity boom type stackers mounted on rails are provided to pile the coal being received.

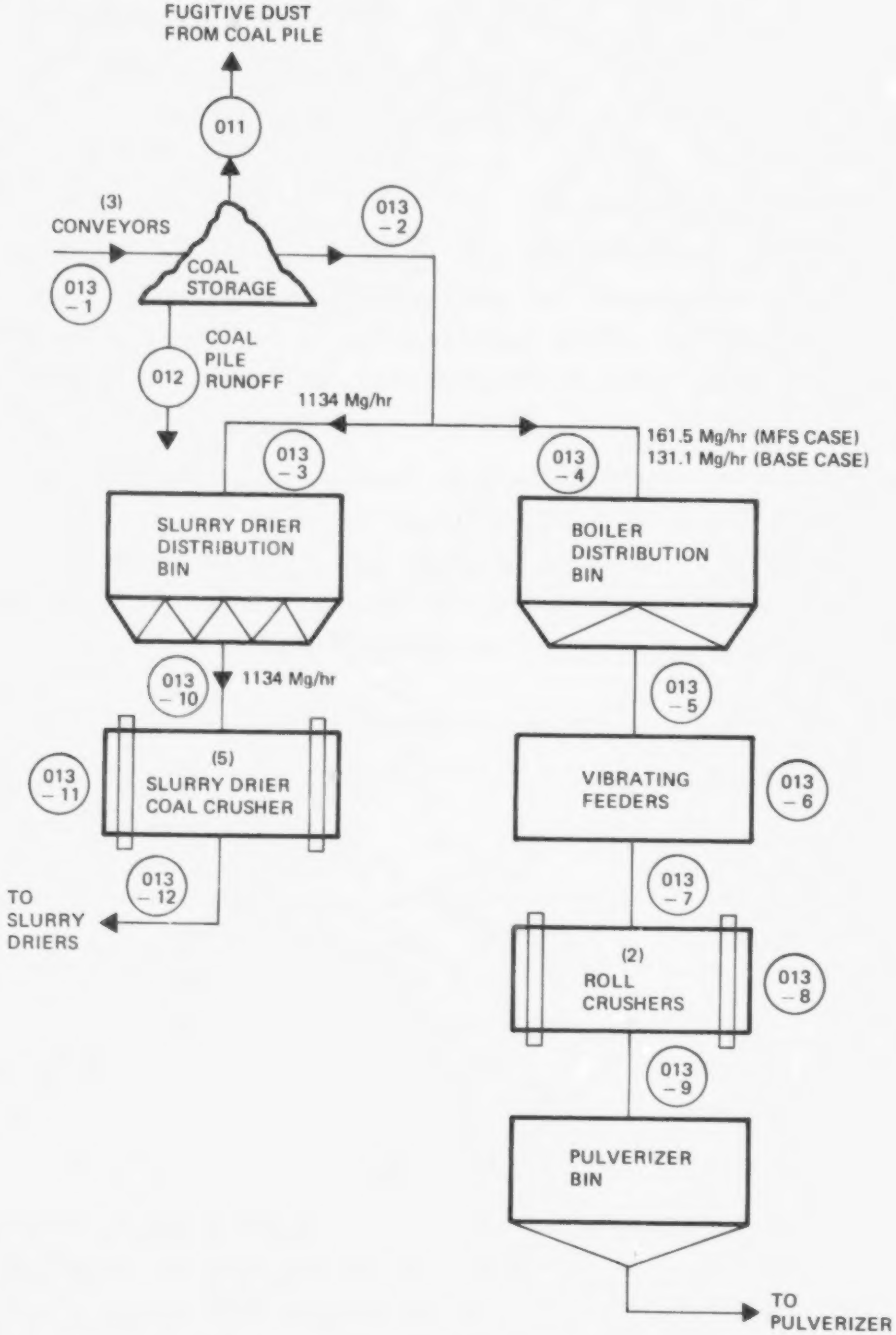
FUGITIVE DUST FROM COAL PILE
011
(3) CONVEYORS
013 – 1
COAL STORAGE
013 – 2
COAL PILE RUNOFF
012
1134 Mg/hr
161.5 Mg/hr (MFS CASE)
131.1 Mg/hr (BASE CASE)
013 – 3
013 – 4
SLURRY DRIER DISTRIBUTION BIN
BOILER DISTRIBUTION BIN
013 – 10
1134 Mg/hr
013 – 5
013 – 11
(5) SLURRY DRIER COAL CRUSHER
VIBRATING FEEDERS
013 – 6
013 – 12
TO SLURRY DRIERS
013 – 7
(2) ROLL CRUSHERS
013 – 8
013 – 9
PULVERIZER BIN
TO PULVERIZER

Figure 3-3. Block flow diagram for EDS commercial plant coal preparation area (Area 1)

Two 100% capacity reclaimers are used to reclaim coal from the live storage piles (24 hrs/day, 7 days/week). Each is a bucketwheel on boom type reclaimer mounted on rails.

3.3.1.1.1 Gaseous Waste Stream

There is only one air emission stream from coal storage:

- Stream 011 - fugitive dust from coal pile

Fugitive dust emissions from coal pile (Stream 011) (Coal Storage - Area 1)

The quantities of fugitive particulates generated by the 5-day and 30-day storage piles were estimated using methodology described by Jutze et al (15). The four major sources of fugitive particulate emissions are: loading onto piles; equipment and vehicle movement in storage area; wind erosion; and load-out from piles. Emission rates from these sources are dependent on the turnover rate for the pile, methods for adding and removing material, and the pile configuration.

Fugitive emission estimates were calculated using several different formulas (15). These formulas include correction factors which account for such parameters as activity on and around storage piles, silt content of material stockpiled, duration of storage and average surface moisture in different geographic areas. The formula factors used in developing fugitive emission estimates assumed that no control techniques are applied (e.g., application of wetting or crusting agents). Estimated uncontrolled fugitive dust emissions from the 5-day live coal storage piles (Stream 011-1) and the 30-day dead coal storage pile (Stream 011-2) are presented in Tables 3-4 and 3-5, respectively. Total uncontrolled fugitive dust emissions from coal storage would be obtained by summing the values in these two tables.

3.3.1.1.2 Liquid Waste Stream

There is only one liquid waste stream from coal storage:

- Stream 012 - coal pile runoff.

TABLE 3-4. UNCONTROLLED FUGITIVE DUST GENERATION
5-DAY COAL STORAGE PILES (STREAM 011-1)

Activity	Illinois No. 6 Coal	
	Base Case, kg/hr	MFS Case kg/hr
Wind Erosion*	2.22	2.27
Loading On†	7.31	7.59
Loading Off‡	9.39	9.61
Vehicular Activity**	0.25	0.25
TOTAL	19.20	19.72

* Based on a respirable emission factor of 6.4 mg of dust per kg/yr of coal stored (Ref. 16). Since respirable emissions represent only 5 percent of total particulate emissions, a correction of 20 has been applied.

† Emission factor formula used is $\frac{(0.02)(K_1)\ (S_1/1.5)}{(PE/100)^2}$ kg/Mg of material transferred (15).

where K_1 (Activity Factor) = 0.75

S_1 (Silt Content) = 0.5, and

PE (Thornthwaite's precipitation-evaporation index) is 93 for Southwestern Illinois.

‡ Emission factor formula used is $\frac{(0.025)(K_1)\ (S_1/1.5)}{(PE/100)^2}$ kg/Mg of material transferred (15).

where K_1 = 0.77

S_1 = 0.5, and

PE = 93 for Southwestern Illinois

**Emission factor formula used is $\frac{(0.065)(K_1)\ (S_1/1.5)}{(PE/100)^2}$ kg/Mg-yr of material stored (15).

where K_1 = 0.5

S_1 = 0.5

PE = 93 for Southwestern Illinois

TABLE 3-5. UNCONTROLLED FUGITIVE DUST GENERATION 30-DAY COAL STORAGE PILES (STREAM 011-2)

Activity	Illinois No. 6 Coal	
	Base Case, kg/hr	MFS Case, kg/hr
Wind Erosion*	13.31	13.62
Loading On†	7.31	7.59
Loading Off ‡	9.39	9.61
Vehicular Activity**	1.44	1.49
Total	31.45	32.31

* Based on a respirable emission factor of 6.4 mg of dust per kg/yr of coal stored (Ref. 16). Since respirable emissions represent only 5 percent of total particulate emissions, a correction of 20 has been applied.

† Emission factor formula used is $\dfrac{(0.02)(K_1)\ (S_1/1.5)}{(PE/100)^2}$ kg/Mg of material transferred (15)

where K_1 (Activity Factor) = 0.75
S_1 (Silt Content) = 0.5, and
PE (Thornthwaite's precipitation-evaporation index) is 93 for Southwestern Illinois

‡ Emission factor formula used is $\dfrac{(0.025)(K_1)\ (S_1/1.5)}{(PE/100)^2}$ kg/Mg of material transferred (15)

where K_1 = 0.77
S_1 = 0.5, and
PE = 93 for Southwestern Illinois

** Emission factor formula used is $\dfrac{(0.065)(K_1)\ (S_1/1.5)}{(PE/100)^2}$ kg/Mg-yr of material stored (15)

where K_1 = 0.5
S_1 = 0.5
PE = 93 for Southwestern Illinois

Coal pile runoff (Stream 012) (Coal Storage - Area 1)

Only a limited amount of data are available in the published literature on the chemical characteristics of coal pile runoff. However, effluents from some coal storage piles comprised of high sulfur coal have been analyzed. In addition, laboratory leaching tests done at Los Alamos Scientific Laboratories have shown that types and quantities of pollutants released from coal storage piles are similar to those from coal refuse (17).

In general, coal from eastern sources has been found to produce a highly acidic runoff stream with pH's ranging from 2.2 to 3.1. Total suspended solids concentrations are generally low during base flow periods but increase dramatically during storm runoff to levels as high as 2300 mg/l. Sulfate concentrations are also quite high with ranges from 1800 to 9600 mg/l. Concentration of iron and manganese are both very high, ranging from 23 to 1800 mg/l and from 1.8 to 45 mg/l, respectively. Other elements of potential concern include aluminum, mercury, arsenic, and zinc.

Estimates of the amount of runoff produced at the coal storage sites as a result of a 10-year, 24-hour storm were determined as follows. The surface area covered by the 5-day and 30-day storage piles were calculated based on a pile height of 18.3 m (60 ft) and a bulk density of 1.153 Mg/m^3 (72 lb/cu ft.). The storage pile was assumed to be a truncated cone with an angle of repose of 40^o. Annual and 10-year, 24-hour storm rainfall values for Southwestern Illinois were obtained from the Rainfall Frequency Atlas of the U.S.

According to Cox et al. (17), 70 percent of the rainfall incident upon a coal pile ultimately appears as runoff. This factor was therefore used to calculate the total runoff from the coal pile itself. In addition, to allow for the fact that the drainage ditch would not be contiguous to the storage pile, an arbitrary 20 percent of the above areas was used to take this into account. The "rational method" presented in Clark, et al (18), was used to calculate peak runoff flow rates. The runoff coefficient used in the rational method was the ratio of peak runoff rate to average rainfall rate, but is also, by virtue of the units used in the equation, the ratio of runoff quantity to rainfall quantity. While this was a peak rather than average value, it was used in the absence of a better method to calculate the maximum runoff that would be expected from a particular site. A value of 0.17, corresponding to flat lawns with heavy soil, was chosen for all site. The rational formula relates runoff to rainfall in the following manner:

$$Q = cia$$

where Q = Peak runoff rate, cfs
c = runoff coefficient
i = average rainfall intensity, in/hr
a = drainage area, acres.

Estimated average and 10-year, 24-hour storm runoffs from the 5-day live storage piles (Stream 012-1) and 30-day dead storage pile (Stream 012-2) are presented in Table 3-6. Total runoff would be the sum of the runoff values for the 5-day and 30-day piles.

3.3.1.1.3 Solid Waste Stream

There is no solid waste stream from coal storage.

3.3.1.2 Coal Handling (Coal Preparation - Area 1)

A spared system of belt conveyors is used to transport reclaimed coal to the distribution bins (13). The slurry drier distribution bin is provided to distribute coal to five crushers through five gravimetric weight belt feeders. Impact cage mill crushers are used to reduce the reclaimed coal to minus 8 mesh size. A single conveyor conveys the coal from each crusher to the top of one of the five slurry driers.

TABLE 3-6. AVERAGE AND 10-YEAR, 24-HOUR STORM RUNOFF FROM 5-DAY WET STORAGE PILES AT THE EDS PLANT (STREAM 012-1)

Coal Type	Base Case	MFS Case
Quantity of coal in* 5 day storage (Mg)	151,800	155,460
Storage pile volume† (m^3)	131,700	144,600
Ground area covered‡ by storage pile (m^2)	10,630	11,610
10-year, 24-hour storm** runoff (m^3)	990	1,080
Average daily runoff†† (m^3)	70	78

FROM 30-DAY DEAD STORAGE PILE AT THE EDS PLANT (STREAM 012-2)

Coal Type	Base Case	MFS Case
Quantity of coal in* 30 day storage (Mg)	910,700	932,800
Storage pile volume† (m^3)	789,900	809,000
Ground area covered‡ by storage pile (m^2)	52,000	52,400
10-year, 24-hour** storm runoff (m^3)	4,860	4,880
Average daily runoff†† (m^3)	346	347

* Based on 24 hours per day

† Based on 1.153 Mg/m^3 bulk density

‡ Based on a covered radius of 76 m.

** 10-year, 24-hour rainfall in Southwestern Illinois was assumed to be 5 inches

†† Annual rainfall in Southwestern Illinois was assumed to be 40 inches.

A second distribution bin for the steam and power boilers is provided to distribute coal to the pulverizer coal crushers through vibrating feeders. Two roll crushers are used to reduce the coal to minus 2 inch size, which are then conveyed to bins for feed to the pulverizers.

3.3.1.2.1 Gaseous Waste Streams

Fugitive particulates are generated from a variety of sub-sources during the conveying/transfer, screening, and crushing of coal as shown in Figure 3-3. Fugutive particulate emissions from coal pulverizers have been reported to be negligible and not considered here (41). The combined emissions from these sub-sources are denoted as:

- Stream 013 - fugitive particulates from coal handling and crushing

Fugitive particulates from coal handling and crushing (Stream 013) (Coal Handling - Area 1)

Coal crushing/screening data are available from surface coal mining and ore mining operations, the crushed stone, and the manufacturing of coke industries. Published data on particulate emission factors for coal operations are limited. The uncontrolled emission factors that are available vary over a broad range, reflecting the uncertainties in the determination of these factors. One study lists an average uncontrolled emission factor of 0.2 kg/Mg of coal processed for loading and unloading activities in all modes of transport (15) while another study (19) lists emission factors of 0.05 kg/Mg and 0.01 kg/Mg of coal mined for coal loading and unloading operations, respectively (19). The emission factors used in deriving uncontrolled particulate emission rates presented in Table 3-7 are intended to provide approximate emissions generated during the specific operations. The streams associated with the estimated emission values are shown in Figure 3-3. Because of the uncertainties in emission factors, estimated uncontrolled emission rates are given in Table 3-7 as ranges, with the range encompassing the low end to the high end of emission factor estimates (0.02 kg/Mg to o.48 kg/Mg in most cases). For the base case

TABLE 3-7. FUGITIVE PARTICULATE EMISSIONS FROM COAL HANDLING AND CRUSHING (STREAM 013)

	Illinois No. 6		Base Case		MFS Case	
Stream No.	Particulate Emission Source	Estimated Average Emission Factors Uncontrolled kg/Mg	Feed Rate (as received basis) Mg/hr	Uncontrolled Emission Rates kg/hr	Feed Rate (as received basis) Mg/hr	Uncontrolled Emission Rates kg/hr
013-1	Conveying/Transfer	0.02-0.48*	1265.1	25.3-607.2	1295.5	25.9-621.8
013-2, 3, 4	Conveying/Transfer	0.02-0.48*	1265.1	25.3-607.2	1295.5	25.9-621.8
013-5	Conveying/Transfer	0.02-0.48*	131.1	3.0-62.9	161.5	3.2-77.5
013-6	Screening	0.04†	131.1	5.2	161.5	6.5
013-7	Conveying/Transfer	0.02-0.48*	131.1	3.0-62.9	161.5	3.2-77.5
013-8	Crushing	0.04†	131.1	5.2	161.5	6.5
013-9	Conveying/Transfer	0.02-0.48*	131.1	3.0-62.9	161.5	3.2-77.5
013-10	Conveying/Transfer	0.02-0.48*	1134	22.7-544.3	1134	22.7-544.3
013-11	Crushing	0.04†	1134	45.4	1134	45.4
013-12	Conveying/Transfer	0.02-0.48*	1134	22.7-544.3	1134	22.7-544.3
013	Total			161-2548		165-2623

* **Both the low end and the high end of the emission factor range are obtained from Reference 19.** This range covers results from a variety of conveying and transferring applications; some of these applications might have included some degree of fugitive particulate control, creating the low end of the presented range. Because of the uncertainty over the degree of control reflected in the lower end of the range, the full range is presented here, although the intent of the table is to present uncontrolled emissions.

† Emission factor as reported for combined "secondary crushing/screening operations" are 0.08 kg/Mg of coal processed. Arbitrarily assumed contribution from crushing = that from screening = 0.04 kg/Mg of coal processed (19). Preliminary data from Exxon pilot plant on emissions from the impact mill crushing operation indicate a lower emission factor of 0.01 kg/Mg (22).

Emissions from coal pulverizer are negligible (41).

and MFS case designs, uncontrolled fugitive particulate emissions from coal handling and crushing were estimated to be 161-2548 kg/hr and 165-2623 kg/hr, respectively.

In the absence of more specific data several simplifying assumptions were made. No modification to the emission factor was made to correct for the moisture content of the coals. Nonparticulate emissions in the pulverizer/dryer off-gases were assumed to be negligible.

3.3.1.2.2 Liquid Waste Stream

There is no liquid waste stream from coal handling.

3.3.1.2.3 Solid Waste Stream

There is no solid waste stream from coal handling.

3.3.2 Coal Liquefaction Area (Area 2)

In the coal liquefaction area for the EDS commercial plant, crushed coal from coal preparation facilities is fed to the slurry drier, where it is mixed with recycle donor solvent from solvent hydrogenation (13). The dry slurry is pumped to reaction pressure, preheated, mixed with preheated hydrogen, and heated in the slurry preheat furnace to the reaction temperature. The preheated dry slurry is then sent to the liquefaction reactor. The reactor effluent is separated into a recycle gas and a slurry stream. The recycle gas is treated, combined with makeup hydrogen, compressed, preheated, and sent to the slurry preheat furnace. The slurry is let down to nearly atmospheric pressure and sent to the product fractionation facilities. A block flow diagram indicating all the major process and waste streams for the coal liquefaction area is presented in Figure 3-4.

3.3.2.1 Slurry Drying and Liquefaction (Coal Liquefaction - Area 2)

In the slurry dreiers, crushed coal feed (minus 8 mesh) is mixed with recycle solvent to form a coal slurry (13). The slurry driers are pressure vessels where the heat for drying is provided by the hot recycle solvent introduced at 558^{o}K (545^{o}F). The operating temperature of the slurry driers is 408^{o}K (275^{o}F). Concurrently with the mixing process, the slurry is dried to less than 4 wt % moisture on dry coal feed basis. The dry product slurry is then pumped to reaction pressure, further heated by heat exchange with liquefaction reactor effluents, and fed to the liquefaction slurry preheat furnace. The overhead from the slurry drier consisting of water vapor (dirven from the coal), stripped recycle donor solvent and some coal volatiles is condensed, cooled and separated into phenolic water, distillate liquid, and offgas. The phenolic wastewater (Stream 103) is withdrawn from the drum and sent to the phenol extraction unit while the distillate is returned to the slurry drier. The slurry drier vent gas (Stream 102) is burned as fuel.

The coal slurry from the slurry drier is mixed with preheated hydrogen

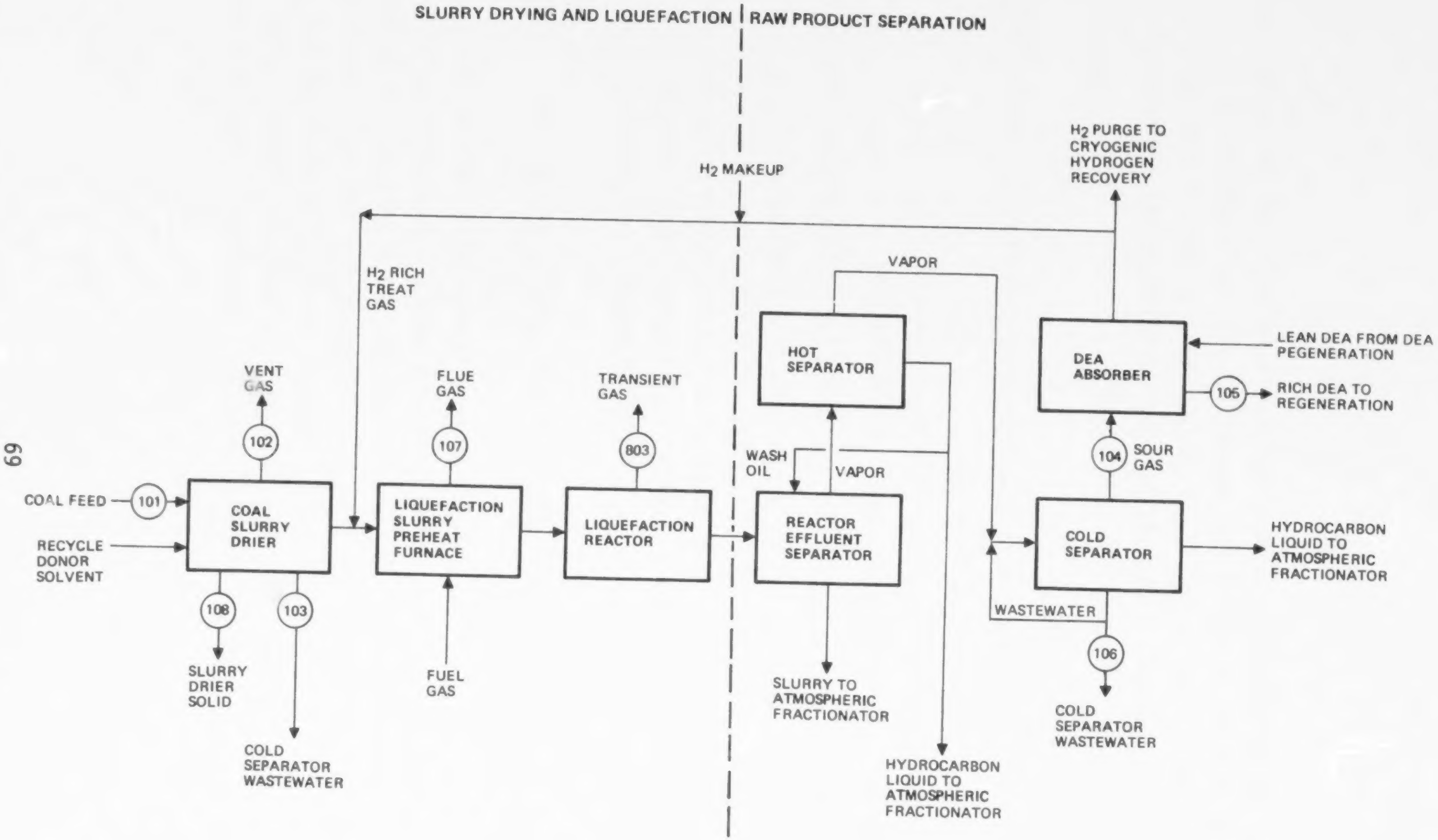
SLURRY DRYING AND LIQUEFACTION
RAW PRODUCT SEPARATION
H_2 MAKEUP
H_2 PURGE TO CRYOGENIC HYDROGEN RECOVERY
H_2 RICH TREAT GAS
VAPOR
VENT GAS
FLUE GAS
TRANSIENT GAS
HOT SEPARATOR
DEA ABSORBER
LEAN DEA FROM DEA PEGENERATION
105
RICH DEA TO REGENERATION
102
107
803
WASH OIL
VAPOR
104
SOUR GAS
COAL FEED
101
RECYCLE DONOR SOLVENT
COAL SLURRY DRIER
LIQUEFACTION SLURRY PREHEAT FURNACE
LIQUEFACTION REACTOR
REACTOR EFFLUENT SEPARATOR
COLD SEPARATOR
HYDROCARBON LIQUID TO ATMOSPHERIC FRACTIONATOR
108
103
WASTEWATER
106
SLURRY DRIER SOLID
FUEL GAS
SLURRY TO ATMOSPHERIC FRACTIONATOR
COLD SEPARATOR WASTEWATER
COLD SEPARATOR WASTEWATER
HYDROCARBON LIQUID TO ATMOSPHERIC FRACTIONATOR

Figure 3-4. Block flow diagram for EDS commercial plant coal liquefaction area (Area 2)

treat gas and fed to the slurry preheat furnace fired with fuel gas. The slurry-hydrogen mixture is heated to liquefaction reactor conditions of 700-750°K (800-900°F) and 13.9 MPa (2000 psig). This mixture is then sent to the liquefaction reactor, where the coal is liquefied in the presence of molecular hydrogen and the hydrogen donor solvent.

3.3.2.1.1 Gaseous Waste Streams

There are three gaseous waste streams from slurry drying and liquefaction:

- Stream 102 - slurry drier vent gas
- Stream 107 - flue gas from slurry preheat furnace
- Stream 803 - transient waste gas from liquefaction.

Slurry drier vent gas (Stream 102) (Slurry drying and liquefaction - Area 2)

The slurry drier vent gas is the offgas from the condensation of the slurry drier overhead vapor. This stream was estimated to be generated at the rate of 532.1 kmol/hr (20). The composition of the slurry drier vent gas is given as follows (6): 86.0% N_2, 9.9% H_2O, 4.1% hydrocarbons (average molecular weight of 97.6, mostly in the C_6 - 400°F normal boiling range), and 30 ppmv H_2S.

Flue gas from slurry preheat furnace (Stream 107) (Slurry drying and liquefaction - Area 2)

This flue gas stream is generated from the preheating of the coal feed slurry by the combustion of the low-Btu fuel gas obtained from the Flexicoking operation. Total heat provided by the fuel gas is 1.885 TJ/hr (1785.6 MM Btu/hr). The low-Btu fuel gas contains 80 ppmv H_2S and 3 ppmv COS (13). Assuming combustion with 20% excess air, the flow rate of the flue gas generated was calculated to be 34,535 kmol/hr (480,920 SCFM). In Table 3-8, the emission factors used and the estimated emission rates of pollutants for this flue gas stream are presented.

TABLE 3-8. ESTIMATED UNCONTROLLED EMISSIONS FOR FLUE GAS FROM SLURRY PREHEAT FURNACE (STREAM 107)

Pollutant	Emission Factor* ng/J (lb/MM Btu)	Emission Rate kg/hr
Particulate Matter	2.1 - 6.4 (0.005-0.015)	4.1 - 12.1
NO_x (as NO_2)	52 - 99 (0.12-0.23)	97 - 186
CO	7.3 -10.7 (0.017-0.025)	14 - 20
Hydrocarbons	1.3 (0.003)	2.4
SO_2	48 (0.11)	91

* Emission factors for particulate matter, NO_x, and hydrocarbons are obtained from AP-42 for natural gas combustion in industrial process heaters (21). By comparison, Exxon estimated emission factors of 3.1 ng/J for particulate matter, and 60 ng/J for NO_x (20). The emission factor for CO is based on 7.3 ng/J from AP-42 and 10.7 ng/J from Exxon. The emission factor for SO_2 is based on 83 ppmv total reduced sulfur in the low-Btu fuel gas (13).

Transient waste gas from liquefaction (Stream 803) (Slurry drying and liquefaction - Area 2)

Gas flows in the liquefaction reactors during startup, shutdown, or upset conditions are considered to be transient waste gases because they may not be suitable for use within the liquefaction facility. For the four liquefaction reactors, it is assumed that transient conditions would be expected twice per year per reactor, and up to 3 hours in duration each time. Also, outage for the reactors will not occur simultaneously. The transient waste gas from the liquefactor reactor is anticipated to be similar in composition to the liquefaction separator sour gas (Stream 104), assuming that the vapor-liquid separation stages are operational under transient conditions. Characterization data for this transient waste gas stream are presented in Table 3-9. The estimated flow rate of this stream is 3,280 kmol/hr (20).

3.3.2.1.2 Liquid Waste Stream

There is only one liquid waste stream from slurry drying and liquefaction:

- Stream 103 - slurry drier cold separator wastewater

Slurry drier cold separator wastewater (Stream 103) (Slurry drying and liquefaction - Area 2)

The estimated characteristics of this wastewater stream are presented in Table 3-10. These characteristics were estimated by Exxon based on analyses of wastewater samples obtained from several small operating units together with computer process synthesis (13,20). Stream 103 is a non-sour phenolic wastewater stream that can be sent directly to the phenol extraction unit without sour water stripping.

3.3.2.1.3 Solid Waste Stream

There is only one solid waste stream from slurry drying and liquefaction:

- Stream 108 - solids accumulated in the slurry drier.

TABLE 3-9. ESTIMATED CHARACTERISTICS OF TRANSIENT WASTE GAS FROM LIQUEFACTION REACTORS (STREAM 803)

Component	Volume Percent in Waste Gas	Flow Rate, kmol/hr
H_2	50.5	1657.4
C_1	34.9	1144.0
C_2	6.6	217.8
C_3	2.5	81.2
C_4	0.7	22.5
C_5	0.2	5.7
C_6 - 400°F	0.2	6.6
400/700°F	20 ppmv	0.07
CO	0.5	15.2
CO_2	1.2	38.4
H_2S	2.5	83.6
NH_3	49 ppmv	0.16
H_2O	0.07	2.3
Total flow rate		3275
Temperature, °K		316
Pressure, MPa		12.7

Data source: Characteristics of Stream 104 in Attachment 3 of Reference 20.

Average frequency of discharge: Two occasions per year for each of 4 reactors, up to 3 hours in duration per occasion.

Note: The single numbers presented in this table represent a specific set of design/operating conditions for the EDS plant. This composition could vary in practice due to uncertainty in the emission estimates, and due to variations in coal type, plant design conditions and plant operating conditions. This note is equally applicable to other composition tables throughout Section 3.

TABLE 3-10. ESTIMATED QUALITY OF SLURRY DRIER COLD SEPARATOR WASTEWATER (STREAM 103)

Contaminant	Concentration ppmw	Flow Rate kg/hr
H_2S	4	0.6
NH_3	76	12
HCl	13	2
CO_2	290	46
Phenols*	970	150
Total flow rate, kg/hr		155,640
Total flow rate, m^3/hr		157.1
Temperature, °K		317
Pressure, MPa		0.09

Data source: Table 1-X-1 of Reference 13.

* Organics other than phenols may also be present. However, no data on the identity or concentration levels of other organics are publicly available.

Solids accumulated in the slurry drier (Stream 108) (Slurry drying and liquefaction - Area 2)

Coal feed to the EDS process has a nominal top-size of 8 mesh (2.38 mm). Oversize feed is intermittently removed from the slurry drier for disposal (Stream 108). A removal frequency of one eight-hour shift every three months has been estimated. Approximately 204 Mg are recovered from the slurry drier during each removal operation (6). The corresponding annual generation rate for this waste is 820 Mg/yr. The slurry drier solids are similar in composition to the feed coal.

3.3.2.2 Raw Product Separation (Coal Liquefaction - Area 2)

In the raw product separation section, the liquefaction reactor effluent is being separated into liquids for further fractionation, and light gases (after acid gas removal) for plant use as recycle hydrogen-rich gas to the liquefaction reactor.

As shown in Figure 3-4, the reactor effluent is separated into a vapor stream and a slurry stream in a vapor-liquid separator. Wash oil from the hot separator drum is used to minimize solids entrainment in the overhead vapor. The vapor from the vapor-liquid separator is cooled by heat exchange and enters the hot separator drum. The hot separator vapor flows to a venturi mixer, where cold separator water is added to prevent ammonia chloride deposition. This mixture is then cooled by additional heat exchange and fed to the cold separator drum. The water added to the hot separator vapor in the mixer is condensed and removed in the cold separator drum, removing, along with it, the ammonia formed in the process.

The vapor from the cold separator, now consisting largely of hydrogen, light hydrocarbons and acid gases, is scrubbed with DEA for acid gas removal. A portion of the resulting hydrogen-rich gas is directed to the cryogenic hydrogen concentration section. The remaining hydrogen-rich gas is mixed with makeup hydrogen and hydrogen purge from the solvent hydrogenation section to form recycle treat gas. The treat gas is compressed, preheated by heat exchange with effluent separator vapor, and recycled back to the slurry furnace.

The slurry stream from the reactor effluent separator is let down to nearly atmospheric pressure and sent directly to the atmospheric fractionator. A portion of the liquid stream from the hot separator is recycled to the reactor effluent separator as wash oil. The remainder of the hot separator liquid is let down in pressure and fed to the atmospheric fractionator. The cold separator hydrocarbon liquid is let down in pressure, preheated by heat exchange with hot separator vapor, and fed to the atmospheric fractionator. A portion of the cold separator sour water is recycled and mixed with hot separator vapor to prevent ammonium chloride plugging. The remainder of the liquefaction cold separator sour water (Stream 106) is sent to the sour water stripper/ammonia recovery system for treatment.

3.3.2.2.1 Gaseous Waste Stream

There is no gaseous waste stream from raw product separation.

3.3.2.2.2 Liquid Waste Stream

There is only one liquid waste stream from raw product separation:

- Stream 106 - Liquefaction cold separator wastewater.

Liquefaction cold separator wastewater (Stream 106) (Raw product separation - Area 2)

The liquefaction cold separator wastewater is generated from the condensation of the water vapor present in the liquefaction reactor effluent. The estimated characteristics of this wastewater stream are presented in Table 3-11. Three sets of values are presented. The first set of values are design estimates developed by Exxon using analyses of wastewater samples from small operating units together with computer process synthesis (13,20), prior to the availability of the data from the large pilot plant. The second and third sets of values are respectively results from EPA (37) and Exxon (23) test campaigns on the 250 ton coal/day EDS pilot plant at Baytown, Texas. As noted in Table

TABLE 3-11. CHARACTERISTICS OF LIQUEFACTION COLD SEPARATOR WASTEWATER (STREAM 106) - THREE DATA SOURCES

Contaminant	Exxon Design Estimate	EPA Testing of EDS Pilot Plant	Exxon Testing of EDS Pilot Plant
H_2S	32,330 ppmw	5,793 mg/l	5,527 mg/l
NH_3	22,840 ppmw	6,910 mg/l	8,400 mg/l
HCl	2,470 ppmw	47 mg/l	No data
CO_2	19,450 ppmw	No data	No data
Phenols	13,720 ppmw	16,024 mg/l	3,650 mg/l*
Organic Acids	7,410 ppmw	No data	No data
Sulfate	No data	36 mg/l	No data
Cl^-	No data	46 mg/l	No data
F^-	No data	30 mg/l	3 mg/l
NO_3^-	No data	2.25 mg/l	No data
NO_2^-	No data	0.02 mg/l	No data
SCN^-	No data	1,800 mg/l	3 mg/l
COD	No data	49,175 mg/l	66,450 mg/l
TDS	No data	11,514 mg/l	8,245 mg/l
TSS	No data	25 mg/l	4 mg/l
TOC	No data	23,512 mg/l	14,500 mg/l
Alkalinity (mg/l as $CaCO_3$)	No data	24,618 mg/l	30,500 mg/l
Oil and Grease	No data	<20 mg/l	No data
Total N	No data	7,294 mg/l	8,500 mg/l
Total S	No data	1,942 mg/l	5,200 mg/l
Flow rate, kg/hr	100,583	No data	No data
Flow rate, m^3/hr	102.6	No data	No data
Temperature °K	317	No data	No data
Pressure, MPa	0.09	No data	No data

Data source: Table 1-X-1 of Reference 13 for Exxon design estimate, Table 6-8 of Reference 37 for EPA testing results and Reference 23 for Exxon testing results.

*GC/MS results by Exxon indicate that this number is too low.

3-11, there are significant differences between the three sets of values presented. Concentration values for the Exxon design estimate are generally greater than those reported by EPA and Exxon test results. A partial explanation may be that less water is condensed in the cold separator for the Exxon commercial plant design compared to the pilot plant operation. According to Exxon, three times more water (per unit coal input basis) may be condensed in the EDS pilot plant as in the Exxon design estimate (22). Also, the elevated level of thiocyanates and the lower level of total sulfur in EPA test results indicate that some of the sulfide present could have participated in the formation of thiocyanates in the sample collected. For purposes of evaluating wastewater treatment technology, the Exxon design estimate values will be used wherever available, even though the design estimates are based upon data from experimental equipment that is much smaller in scale than the Baytown pilot plant. This decision to use the design values is made because the two sets of pilot plant data have not yet been correlated to process conditions, or incorporated into the material balance for the overall plant by Exxon. These pilot plant data, if used independently without proper integration, may result in erroneous wastewater balances for the EDS commercial plant based on the base case and MFS case designs. The Exxon design estimate values will be supplemented by values from the Exxon and EPA test results where necessary. In providing these supplemental values, Exxon test results are preferred over EPA test results because unlike the EPA samples which necessarily had to be shipped for analysis to a laboratory outside the Baytown site, Exxon samples could be analyzed on site minimizing sample degradation. Accordingly, the characteristics of Stream 106 selected for use in the PCTM wastewater treatment technology evaluation are summarized in Table 3-12. For each species, the table indicates whether design estimates, pilot data, or some combination of design estimates and pilot data were used to arrive at the concentration for that particular species.

In Table 3-13, the concentrations of trace metals in Stream 106, as determined by EPA and Exxon from pilot plant wastewater samples, are presented. As

TABLE 3-12. CHARACTERISTICS OF LIQUEFACTION COLD SEPARATOR WASTEWATER USED FOR PCTM IN WASTEWATER TREATMENT TECHNOLOGY EVALUATION (STREAM 106)

Contaminant	Concentration ppmw*	Flow Rate, kg/hr	Source for Selection of Conc. Value
H_2S	32,330	3,251	Design
NH_3	22,840	2,297	Design
HCl	2,470	249	Design
CO_2	19,450	1,956	Design
Phenols	13,720	1,380	Design
Organic Acids	7,410	745	Design
Sulfate	37	3.7	EPA Data
F^-	3	0.3	Exxon Data
NO_3^-	2	0.2	EPA Data
NO_2^-	0.02	0.002	EPA Data
SCN^-	3	0.3	Exxon Data
COD	58,970	5,933	Avg. Exxon/EPA Data
TDS	8,410	810	Exxon Data
TSS	4	0.4	Exxon Data
TOC	19,390	1,950	Avg. Exxon/EPA Data
Alkalinity (ppmw as $CaCO_3$)	31,100	3,130	Exxon Data
Total flow rate, kg/hr		100,583	
Total flow rate, m^3/hr		102.6	
Temperature, °K		317	
Pressure, MPa		0.09	

* Note that mg/l values in Table 3-11 have been converted to ppmw values by using the density of the wastewater.

TABLE 3-13. CONCENTRATIONS OF TRACE METALS IN LIQUEFACTION COLD SEPARATOR WASTEWATER (STREAM 106)

Element		EPA Testing of EDS Pilot Plant*	Exxon Testing of EDS Pilot Plant†
Aluminum	Al	0.031	3.18
Antimony	Sb	<0.01	(0.0636)
Arsenic	As	0.004	(0.09)
Barium	Ba	0.084	0.264
Beryllium	Be	<0.003	<0.002
Bismuth	Bi	<0.50	--
Boron	B	383	163‡
Cadmium	Cd	<0.025	<0.016
Calcium	Ca	1.79	4.09
Chromium	Cr	<0.03	0.0629
Cobalt	Co	<0.02	0.0451
Copper	Cu	<0.015	0.324
Iron	Fe	1.64	1.16
Lead	Pb	<0.08	0.0999
Magnesium	Mg	0.18	0.52
Manganese	Mn	0.043	0.0293
Mercury	Hg	Not detected	--
Molybdenum	Mo	<0.04	(0.0182)
Nickel	Ni	0.035	0.405
Phosphorus	P	<0.40	(0.142)
Potassium	K	0.60	0.972
Selenium	Se	<0.001	<0.0353
Silicon	Si	9.91	56.4
Silver	Ag	<0.03	<0.0086
Sodium	Na	3.92	7.69
Strontium	Sr	<0.001	0.0146
Tin	Sn	<0.03	0.0954
Titanium	Ti	<0.006	0.041
Tungsten	W	--	<0.101
Uranium	U	--	<0.0487
Vanadium	V	<0.01	0.541
Zinc	Zn	0.086	0.893

Data source: Reference 24 for EPA testing results, Reference 25 for Exxon testing results.

-- Not analyzed

* All EPA values are in units of mg/l

† All Exxon values are in units of ppmw

‡ Value out of range, needs further dilution

( ) Value may not be significant.

can be seen from this table, the EPA and Exxon test results differ significantly, possibly because the samples were taken two weeks apart. Both sets of values, however, indicate low levels of trace metals for this wastewater stream.

In Table 3-14, preliminary results of organic analysis obtained by EPA from the EDS pilot plant wastewater sample are presented. These results provide some insight on the types of organic compounds present, and indicate that the major organic components are aliphatic hydrocarbons and phenols.

The liquefaction cold separator wastewater stream will likely be combined with other process wastewater streams from the product separation and purification and liquefaction residue processing/hydrogen plant process areas, prior to treatment in the sour water stripper/ammonia recovery system.

3.3.2.2.3 Solid Waste Stream

There is no solid waste stream from raw product separation.

TABLE 3-14. CONCENTRATIONS OF ORGANIC COMPOUNDS IN LIQUEFACTION COLD SEPARATOR WASTEWATER (STREAM 106)

Organic Compound Class	Concentration mg/l
Aliphatic Hydrocarbons	2,170
Ethers	260
Aldehydes and Ketones	51
Nitrites	310
Amines	7.4
Thiols, Sulfides and Disulfides	31
Benzene and Substituted Benzene Hydrocarbons	140
Phenols*	2,150
Fused Polycyclic Hydrocarbons†	18
Fused Non-Alternant Polycyclic Hydrocarbons ‡	17
Heterocyclic Nitrogen Compounds	120
Heterocyclic Oxygen Compounds	320
Heterocyclic Sulfur Compounds (Thiophenes)	18
Total	5,610

Data source: Table 6-6 of Reference 37. These results are preliminary.

* Breakdown for the phenols: 1,200 mg/l phenol, 150 mg/l o-cresol, 410 mg/l m-cresol, 150 mg/l p-cresol, 190 mg/l C_2 phenol, 19 mg/l C_3 phenol, 2.3 mg/l C_4 phenol, 14 mg/l methyl naphthol, 15 mg/l C_2 naphthol.

† The only fused polycyclic hydrocarbons identified were naphthalene and methyl naphthalene.

‡ The only fused non-alternant polycyclic hydrocarbons identified were fluorene and methyl fluorene.

3.3.3 Product Separation and Purification Area (Area 3)

The product separation and purification operations for the EDS commercial plant consist of four major process areas: liquefaction product fractionation, solvent hydrogenation, gas treating, and product recovery (13). A block flow diagram indicating all the major process and waste streams from these operations is presented in Figure 3-5.

3.3.3.1 Liquefaction Product Fractionation (Product Separation/Purification - Area 3)

The slurry streams from the liquefaction hot and cold separators (Area 2) are fed to the atmospheric fractionator (or tower) at different points (13). In the atmospheric tower, the reactor effluent is separated into offgas, naphtha, a 400/650°F low sulfur fuel oil (LSFO)/spent solvent sidestream, and a slurry bottoms stream. The 400/650°F LSFO/spent solvent cut is steam stripped in a sidestream stripper tower in order to meet flash point requirements for the LSFO product and for the solvent, and sent to solvent hydrogenation area. The atmospheric tower offgas (Stream 151), a sour gas consisting primarily of hydrogen, light hydrocarbons, and hydrogen sulfide, is sent to the DEA scrubber for acid gas removal. The overhead drum wastewater (Stream 152), resulting from additional condensation of water vapor present in the liquefaction reactor effluent, is sent to wastewater treatment. Naphtha separated from the overhead drum wastewater is sent to the product recovery area for additional processing.

The atmospheric tower bottoms are let down in pressure and fed directly to the vacuum fractionator. Products from the vacuum fractionator include offgas, vacuum distillate, vacuum gas oil (VGO), 650/900°F LSFO/spent solvent sidestreams, and the vacuum bottoms which are fed to the Flexicoker. The vacuum fractionator offgas (Stream 153) is burned as fuel. The sour water separated from the vacuum distillate is combined with the condensate from the

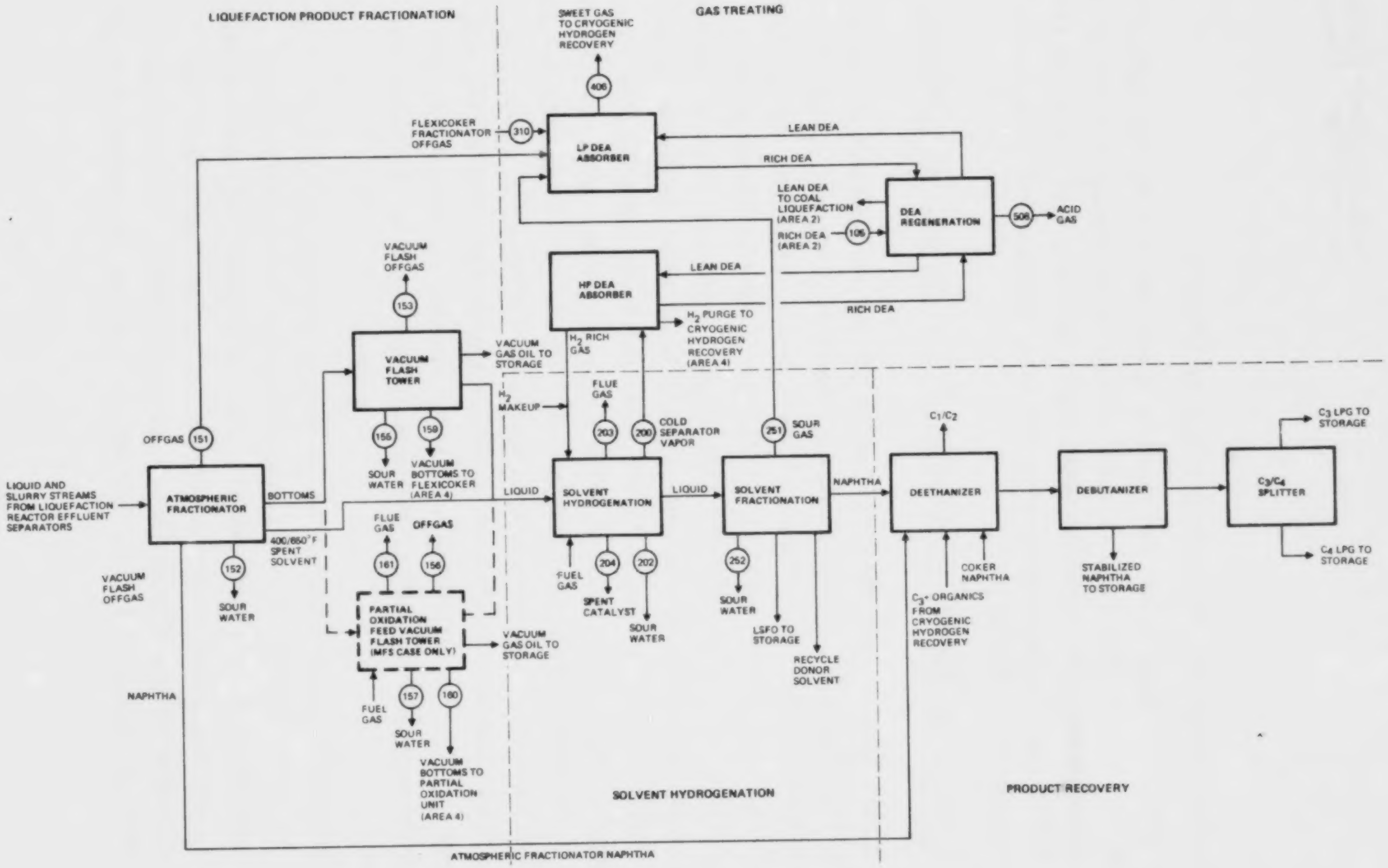
LIQUEFACTION PRODUCT FRACTIONATION
GAS TREATING
SOLVENT HYDROGENATION
PRODUCT RECOVERY
LIQUID AND SLURRY STREAMS FROM LIQUEFACTION REACTOR EFFLUENT SEPARATORS
ATMOSPHERIC FRACTIONATOR
OFFGAS
151
152
SOUR WATER
VACUUM FLASH OFFGAS
NAPHTHA
BOTTOMS
400/850°F SPENT SOLVENT
VACUUM FLASH TOWER
VACUUM FLASH OFFGAS
153
155
SOUR WATER
159
VACUUM BOTTOMS TO FLEXICOKER (AREA 4)
VACUUM GAS OIL TO STORAGE
PARTIAL OXIDATION FEED VACUUM FLASH TOWER (MFS CASE ONLY)
FLUE GAS
161
OFFGAS
156
FUEL GAS
157
SOUR WATER
160
VACUUM BOTTOMS TO PARTIAL OXIDATION UNIT (AREA 4)
VACUUM GAS OIL TO STORAGE
FLEXICOKER FRACTIONATOR OFFGAS
310
LP DEA ABSORBER
406
SWEET GAS TO CRYOGENIC HYDROGEN RECOVERY
LEAN DEA
RICH DEA
DEA REGENERATION
LEAN DEA TO COAL LIQUEFACTION (AREA 2)
RICH DEA (AREA 2)
105
506
ACID GAS
HP DEA ABSORBER
LEAN DEA
RICH DEA
H_2 RICH GAS
H_2 PURGE TO CRYOGENIC HYDROGEN RECOVERY (AREA 4)
H_2 MAKEUP
LIQUID
SOLVENT HYDROGENATION
FLUE GAS
203
200
COLD SEPARATOR VAPOR
FUEL GAS
204
SPENT CATALYST
202
SOUR WATER
LIQUID
SOLVENT FRACTIONATION
251
SOUR GAS
252
SOUR WATER
LSFO TO STORAGE
RECYCLE DONOR SOLVENT
NAPHTHA
DEETHANIZER
C_1/C_2
COKER NAPHTHA
C_3^+ ORGANICS FROM CRYOGENIC HYDROGEN RECOVERY
DEBUTANIZER
STABILIZED NAPHTHA TO STORAGE
C_3/C_4 SPLITTER
C_3 LPG TO STORAGE
C_4 LPG TO STORAGE
ATMOSPHERIC FRACTIONATOR NAPHTHA

Figure 3-5. Block flow diagram for EDS commercial plant product separation and purification area (Area 3)

vacuum jet ejector (Stream 155), and sent to wastewater treatment. The VGO obtained is sent to liquid storage. The vacuum distillate and 650/900°F LSFO/ spent solvent are sent to the solvent hydrogenation area.

For the MFS case, a vacuum fractionator for the partial oxidation feed is added to process part of the atmospheric tower bottoms. The major difference between this fractionator and the Flexicoker feed vacuum fractionator is preheating of the atmospheric tower bottoms in furnaces before feeding to the partial oxidation feed vacuum fractionator. As a result, the partial oxidation feed vacuum fractionator bottoms have a cut point of 975°F vs. the Flexicoker feed vacuum fractionator bottoms cut point of 920°F. The deeper cut point allows the recovery of additional heavy liquids which would be destroyed if left in the bottoms feed to partial oxidation. Similar process and waste streams are generated from the partial oxidation feed vacuum fractionator, including a vacuum fractionator offgas (Stream 156) burned as fuel, a sour water (Stream 157) sent to wastewater treatment, a VGO to liquid storage, a vacuum distillate and a 650/900°F LSFO/spent solvent sent to the solvent hydrogenation area.

3.3.3.1.1 Gaseous Waste Streams

There are three gaseous waste streams from liquefaction product fractionation:

- Stream 153 - offgas from vacuum fractionator feeding Flexicoker
- Stream 156 - offgas from vacuum fractionator feeding partial oxidation unit (MFS case only)
- Stream 161 - flue gas from preheat furnace for vacuum fractionator feeding partial oxidation unit (MFS case only).

Vacuum fractionator offgases (Streams 153 and 156) (Liquefaction product fractionation - Area 3)

The vacuum fractionator offgases result from the letdown in pressure of the atmospheric fractionator bottoms slurry feed to the vacuum fractionator.

For the base case design, the vacuum fractionator offgas (Stream 153) generation rate was estimated to be 68.9 kmol/hr (20). For the MFS case, the combined vacuum fractionator offgas (Stream 153) and partial oxidation feed vacuum fractionator offgas (Stream 156) generation rate was estimated to be 68.9 kmol/hr (20). For both the base case and the MFS case, the offgas is estimated to contain 7.5% O_2, 28.7% N_2, 11.3% H_2O, 47.5% hydrocarbons (with average molecular weight of 46.2), and 5.0% H_2S (8).

Flue gas from preheat furnace for vacuum fractionator feeding partial oxidation unit (Stream 161) (Liquefaction product fractionation - Area 3)

This flue gas stream is generated only in the MFS case design and results from the combustion of the low-Btu fuel gas in the preheater for the atmospheric tower bottoms. The heated atmospheric tower bottoms are sent to the vacuum fractionator designed to produce vacuum bottoms feed for the partial oxidation unit. Total heat provided by the combustion of the low-Btu fuel gas from the Flexicoking unit is 0.023 TJ/hr (22.0 MM Btu/hr). The low-Btu fuel gas contains 80 ppmv H_2S and 3 ppmv COS (13). With 20% excess air, the flow rate of this flue gas stream was calculated to be 425 kmol/hr (5925 SCFM). In Table 3-15, the emission factors used and the estimated emission rates of pollutants for this flue gas are presented.

3.3.3.1.2 Liquid Waste Stream

There are three aqueous waste streams from liquefaction product fractionation:

- Stream 152 - atmospheric fractionator overhead drum wastewater
- Stream 155 - vacuum fractionator wastewater
- Stream 157 - partial oxidation feed vacuum fractionator wastewater (MFS case only).

TABLE 3-15. ESTIMATED UNCONTROLLED EMISSIONS FOR FLUE GAS FROM PARTIAL OXIDATION FEED VACUUM FRACTIONATOR PREHEAT FURNACE (STREAM 161)

Pollutant	Emission Factor* ng/J (lb/MM Btu)	Emission Rate kg/hr
Particulate Matter	2.1 - 6.4 (0.005-0.015)	0.05-0.15
NO_x (as NO_2)	52 - 99 (0.12-0.23)	1.2 -2.3
CO	7.3-10.7(0.017-0.025)	0.17-0.25
Hydrocarbons	1.3 (0.003)	0.03
SO_2	48 (0.11)	1.1

* Emission factors for particulate matter, NO_x, and hydrocarbons are obtained from AP-42 for natural gas combustion in industrial process heaters (21). By comparison, Exxon estimated emission factors of 3.1 ng/J for particulate matter, and 60 ng/J for NO_x (20). The emission factor for CO is based on 7.3 ng/J from AP-42 and 10.7 ng/J from Exxon. The emission factor for SO_2 is based on 83 ppmv total reduced sulfur in the low-Btu fuel gas (13).

Atmospheric fractionator overhead drum wastewater (Stream 152) (Liquefaction product fractionator - Area 3)

The atmospheric fractionator overhead drum wastewater is generated as a result of the additional condensation of water vapor present in the liquefaction reactor effluent, and condensation of steam fed to the atmospheric fractionator sidestream stripper. The estimated characteristics of this wastewater stream are presented in Table 3-16. Three sets of values are presented. The first set of values represents design estimates by Exxon based on analyses of wastewater samples from small operating units together with computer proces synthesis (13,20), prior to the availability of data from the large pilot plant. The second and third sets of values represent EPA and Exxon test results from the 250 ton coal/day EDS pilot plant, respectively. Comparison of these three sets of values shows that H_2S and NH_3 concentrations are considerably higher and the phenolic concentration is considerably lower in the Exxon design estimate. Direct comparisons are difficult because of possible differences in process operating conditions. According to Exxon (22), there may be more than three times the amount of water condensed (per unit coal input basis) in the pilot plant as in the Exxon design estimate. The Exxon design estimates are used for purposes of evaluating wastewater treatment technology wherever possible. This is because the two sets of pilot plant data have not yet been correlated to process conditions, or incorporated into the material balance for the overall plant. These pilot plant data, if used independently without proper integration, may result in erroneous wastewater balances for the EDS commercial plant based on the base case and MFS case designs. The Exxon design estimate values will be supplemented by values from the Exxon and EPA test results as necessary. As discussed for Stream 106 previously, Exxon test results are preferred over EPA test results because Exxon samples were analyzed on site minimizing sample degradation. Accordingly, the characteristics of Stream 152 selected for use in the PCTM wastewater treatment technology evaluation are summarized in Table 3-17. For each species, the table indicates

TABLE 3-16. CHARACTERISTICS OF ATMOSPHERIC FRACTIONATOR OVERHEAD DRUM WASTEWATER (STREAM 152) - THREE DATA SOURCES

Contaminant	Exxon Design Estimate	EPA Testing of EDS Pilot Plant	Exxon Testing of EDS Pilot Plant
H_2S	4,200 ppmw	200 mg/l	372 mg/l
NH_3	5,400 ppmw	1,730 mg/l	1,800 mg/l
HCl	1,270 ppmw	125 mg/l	No data
CO_2	330 ppmw	No data	No data
Phenols	18,400 ppmw	27,861 mg/l	39,500 mg/l
Organic Acids	3,130 ppmw	No data	No data
Sulfate	No data	63 mg/l	No data
Cl^-	No data	122 mg/l	No data
F^-	No data	8 mg/l	1 mg/l
NO_3^-	No data	0.15 mg/l	No data
NO_2^-	No data	Not detected	No data
SCN^-	No data	240 mg/l	2 mg/l
COD	No data	93,700 mg/l	119,500 mg/l
TDS	No data	678 mg/l	175 mg/l
TSS	No data	31 mg/l	4 mg/l
TOC	No data	26,504 mg/l	40,000 mg/l
Alkalinity (mg/l as $CaCO_3$)	No data	5,022 mg/l	5,200 mg/l
Oil and Grease	No data	<20 mg/l	No data
Total N	No data	1,985 mg/l	1,850 mg/l
Total S	No data	1,639 mg/l	350 mg/l
Flow rate, kg/hr	29,105	No data	No data
Flow rate, m^3/hr	30.8	No data	No data
Temperature °K	392°K	No data	No data
Pressure, MPa	0.52	No data	No data

Data source: Table 1-X-1 of Reference 13 for Exxon design estimate, Table 6-20 of Reference 37 for EPA testing results and Reference 23 for Exxon testing results.

TABLE 3-17. CHARACTERISTICS OF ATMOSPHERIC FRACTIONATOR OVERHEAD DRUM WASTEWATER USED FOR PCTM WASTEWATER TREATMENT TECHNOLOGY EVALUATION (STREAM 152)

Contaminant	Concentration, ppmw*	Flow Rate kg/hr	Source for Selection of Conc. Value
H_2S	4,200	122	Design
NH_3	5,400	157	Design
HCl	1,270	36.9	Design
CO_2	330	9.5	Design
Phenols	18,400	536	Design
Organic Acids	3,130	91.0	Design
Sulfate	67	2	EPA Data
F^-	1	0.03	Exxon Data
NO_3^-	0.15	0.004	EPA Data
NO_2^-	Not Detected	0	EPA Data
SCN^-	2	0.06	Exxon Data
COD	112,700	3,280	Avg. Exxon/EPA Data
TDS	185	5.4	Exxon Data
TSS	4	0.1	Exxon Data
TOC	35,150	1,024	Avg. Exxon/EPA Data
Alkalinity (ppmw as $CaCO_3$)	5,500	160	Exxon Data
Total flow rate, kg/hr		29,105	
Total flow rate, m^3/hr		30.8	
Temperature, °K		392	
Pressure, MPa		0.52	

* Note that mg/l values in Table 3-16 have been converted to ppmw values by using the density of the wastewater.

whether design estimates, pilot plant data, or some combination were used to arrive at the concentration for the species.

Table 3-18 presents the concentrations of trace metals in this stream as determined by EPA and Exxon from pilot plant wastewater samples. The EPA and Exxon samples were taken two weeks apart. The test results indicate that concentration values determined by EPA are higher than those by Exxon for certain trace metals, and lower for other trace metals. Both sets of data, however, indicate very low levels of trace metals for this wastewater stream.

In Table 3-19, preliminary results of organic analysis obtained by EPA from the EDS pilot plant wastewater samples are presented (37). These results indicate that most organics are present as phenols.

The atmospheric fractionator overhead drum wastewater will likely be combined with other sour wastewater streams from process areas prior to treatment in the sour water stripper/ammonia recovery system.

Vacuum fractionator wastewater (Streams 155 and 157) (Liquefaction product separation - Area 3)

The vacuum fractionator wastewater is the combined wastewater from two sources: 1) wastewater from the vacuum fractionator distillate drum, and 2) steam ejector condensate from the vacuum fractionator. For the base case, this wastewater stream is identified as Stream 155. For the MFS case, there is an additional wastewater stream from the partial oxidation feed vacuum fractionator. This additional wastewater stream is identified as Stream 157. For the MFS case, the combined Stream 155/157 has the same flow rate and characteristics as the single Stream 155 in the base case design.

The estimated characteristics of the vacuum fractionator wastewater are presented in Table 3-20. The three sets of values presented represent Exxon design estimates, and EPA and Exxon pilot plant test results, respectively. In contrast to the atmospheric fractionator overhead drum wastewater, the Exxon pilot test results show a higher H_2S level and a lower phenol level when

TABLE 3-18. CONCENTRATIONS OF TRACE METALS IN ATMOSPHERIC FRACTIONATOR OVERHEAD DRUM WASTEWATER (STREAM 152)

Element		EPA Testing of EDS Pilot Plant*	Exxon Testing of EDS Pilot Plant†
Aluminum	Al	0.024	0.0564
Antimony	Sb	<0.010	<0.0313
Arsenic	As	0.006	(0.0388)
Barium	Ba	0.021	0.0644
Beryllium	Be	<0.003	<0.0012
Bismuth	Bi	<0.50	--
Boron	B	0.054	0.116
Cadmium	Cd	<0.025	(0.012)
Calcium	Ca	2.61	3.48
Chromium	Cr	<0.03	0.0086
Cobalt	Co	<0.02	0.0183
Copper	Cu	<0.015	0.276
Iron	Fe	1.90	0.20
Lead	Pb	<0.08	<0.0212
Magnesium	Mg	0.085	0.206
Manganese	Mn	0.022	0.018
Mercury	Hg	Not detected	--
Molybdenum	Mo	<0.04	<0.0084
Nickel	Ni	<0.025	0.117
Phosphorus	P	<0.40	<0.0465
Potassium	K	0.16	<0.0202
Selenium	Se	0.034	<0.0212
Silicon	Si	2.24	1.42
Silver	Ag	<0.03	0.0172
Sodium	Na	2.16	1.42
Strontium	Sr	0.004	0.0048
Tin	Sn	<0.03	<0.0162
Titanium	Ti	0.008	<0.0011
Tungsten	W	--	<0.0606
Uranium	U	--	<0.0293
Vanadium	V	<0.01	<0.0092
Zinc	Zn	0.091	0.219

Data source: Reference 24 for EPA testing results, Reference 25 for Exxon testing results.

-- Not analyzed

* All EPA values are in units of mg/l

† All Exxon values are in units of ppmw

( ) Value may not be significant

TABLE 3-19. CONCENTRATIONS OF ORGANIC COMPOUNDS IN ATMOSPHERIC FRACTIONATOR OVERHEAD DRUM WASTEWATER (STREAM 152)

Organic Compound Class	Concentration mg/l
Aliphatic Hydrocarbons	3,750
Ethers	560
Aldehydes and Ketones	210
Carboxylic Acids and Derivatives	120
Nitrites	700
Amines	53
Benzene and Substituted Benzene Hydrocarbons	270
Phenols*	15,800
Heterocyclic Nitrogen Compounds	160
Heterocyclic Oxygen Compounds	22
Heterocyclic Sulfur Compounds (Thiophenes)	56
Total	21,700

Data source: Table 6-18 of Reference 37. These results are preliminary.

* Breakdown for the phenols: 13,000 mg/l phenol, 1,100 mg/l o-cresol, 470 mg/l m-cresol, 1,200 mg/l p-cresol, and 27 mg/l C_2 phenol.

TABLE 3-20. CHARACTERISTICS OF VACUUM FRACTIONATOR WASTEWATER (STREAMS 155/157) - THREE DATA SOURCES

Contaminant	Exxon Design Estimate	EPA Testing of EDS Pilot Plant	Exxon Testing of EDS Pilot Plant
H_2S	8,550 ppmw	15 mg/l	29,229 mg/l
NH_3	37.5 ppmw	171 mg/l	145 mg/l
HCl	18.7 ppmw	<1 mg/l	No data
CO_2	730 ppmw	No data	No data
Phenols	4,910 ppmw	813 mg/l	455 mg/l
Organic Acids	340 ppmw	No data	No data
Sulfate	No data	47 mg/l	No data
Cl^-	No data	<1 mg/l	No data
F^-	No data	<1 mg/l	<1 mg/l
NO_3^-	No data	0.18 mg/l	No data
NO_2^-	No data	Not detected	No data
SCN^-	No data	Not detected	3 mg/l
COD	No data	3,701 mg/l	4,835 mg/l
TDS	No data	402 mg/l	135 mg/l
TSS	No data	28 mg/l	15 mg/l
TOC	No data	1,329 mg/l	1,250 mg/l
Alkalinity (mg/l as $CaCO_3$)	No data	728 mg/l	517 mg/l
Oil and Grease	No data	1,154 mg/l	No data
Total N	No data	540 mg/l	155 mg/l
Total S	No data	473 mg/l	27,500 mg/l
Flow rate, kg/hr	4,839	No data	No data
Flow rate, m^3/hr	4.9	No data	No data
Temperature °K	317	No data	No data
Pressure, MPa	0.74	No data	No data

Data source: Table 1-X-1 of Reference 13 for Exxon design estimate, Table 6-32 of Reference 37 for EPA testing results and Reference 23 for Exxon testing results.

compared with the corresponding Exxon design estimates. The H_2S level from the EPA test results was 15 mg/l and appears to be unreasonably low. A partial explanation for the differences in concentration values is the possible differences in process operating conditions and wastewater flow rates (per unit coal input basis) between the Exxon design and the pilot plant. For purposes of wastewater treatment technology evaluation, the Exxon design estimates will be used wherever available. As discussed for Stream 152 previously, this is primarily because the two sets of pilot plant data have not yet been integrated into the EDS commercial plant design by Exxon. In Table 3-21, the characteristics of Streams 155/157 selected for use in PCTM wastewater treatment technology evaluation are presented. The values presented in this table are primarily Exxon design estimate values, supplemented by values from the Exxon and EPA test results only where necessary. In the latter case, Exxon test results are preferred over EPA test results because Exxon samples could be analyzed on site and hence were less subject to sample degradation problems.

In Table 3-22, the concentrations of trace metals in this stream determined by EPA and Exxon using pilot plant wastewater samples are presented. As with other wastewater streams sampled, the EPA and Exxon samples were taken two weeks apart. For most trace metals, the EPA test results indicate higher concentration levels than the Exxon test results. Both sets of data, however, indicate very low levels of trace metals for this wastewater stream.

In Table 3-23, preliminary results of organic analysis obtained by EPA from the EDS pilot plant wastewater samples are presented. These results provide information on the types of organic compounds present in this wastewater stream.

The vacuum fractionator wastewater will also likely be combined with other sour wastewater streams from process areas prior to treatment. By comparison, this wastewater stream is relatively small and contributes less than 2% of the combined sour wastewater stream.

TABLE 3-21. CHARACTERISTICS OF VACUUM FRACTIONATOR WASTEWATER USED FOR PCTM WASTEWATER TREATMENT TECHNOLOGY EVALUATION (STREAMS 155/157)

Contaminant	Concentration, ppmw*	Flow Rate kg/hr	Source for Selection of Conc. Value
H_2S	8,550	41.4	Design
NH_3	37.5	0.2	Design
HCl	18.7	0.1	Design
CO_2	730	3.5	Design
Phenols	4,910	23.8	Design
Organic Acids	340	1.6	Design
Sulfate	48	0.2	EPA Data
F^-	<1	0.005	Exxon Data
NO_3^-	0.18	0.0009	EPA Data
NO_2^-	Not Detected	0	EPA Data
SCN^-	3	0.01	Exxon Data
COD	4,330	20.9	Avg. Exxon/EPA Data
TDS	137	0.7	Exxon Data
TSS	15	0.07	Exxon Data
TOC	1,310	6.3	Avg. Exxon/EPA Data
Alkalinity (ppmw as $CaCO_3$)	524	2.5	Exxon Data
Total flow rate, kg/hr		4,839	
Total flow rate, m^3/hr		4.9	
Temperature, °K		317	
Pressure, MPa		0.74	

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* Note that mg/l values in Table 3-20 have been converted to ppmv values by using the density of the wastewater.

TABLE 3-22. CONCENTRATIONS OF TRACE METALS IN VACUUM FRACTIONATOR WASTEWATER (STREAMS 155/157)

Element		EPA Testing of EDS Pilot Plant*	Exxon Testing of EDS Pilot Plant†
Aluminum	Al	0.18	0.082
Antimony	Sb	< 0.010	< 0.035
Arsenic	As	0.008	(0.0328)
Barium	Ba	0.0305	0.0153
Beryllium	Be	< 0.003	< 0.0014
Bismuth	Bi	< 0.50	--
Boron	B	0.265	0.387
Cadmium	Cd	< 0.025	< 0.0107
Calcium	Ca	0.89	0.281
Chromium	Cr	< 0.03	0.0164
Cobalt	Co	< 0.02	(0.0099)
Copper	Cu	0.0435	0.108
Iron	Fe	0.47	0.203
Lead	Pb	< 0.08	< 0.0181
Magnesium	Mg	0.13	0.124
Manganese	Mn	0.135	0.0071
Mercury	Hg	Not detected	--
Molybdenum	Mo	< 0.04	(0.0127)
Nickel	Ni	0.042	0.107
Phosphorus	P	< 0.040	< 0.052
Potassium	K	0.285	< 0.0226
Selenium	Se	0.004	< 0.0237
Silicon	Si	6.02	3.14
Silver	Ag	< 0.03	< 0.0058
Sodium	Na	2.57	1.09
Strontium	Sr	0.002	0.0009
Tin	Sn	< 0.03	(0.0248)
Titanium	Ti	0.0135	< 0.0012
Tungsten	W	--	< 0.0678
Uranium	U	--	< 0.0328
Vanadium	V	< 0.01	(0.0112)
Zinc	Zn	0.0205	0.255

Data source: Reference 24 for EPA testing results, Reference 25 for Exxon testing results.

-- Not analyzed

* All EPA values are in units of mg/l

† All Exxon values are in units of ppmw

( ) Value may not be significant

TABLE 3-23. CONCENTRATIONS OF ORGANIC COMPOUNDS IN VACUUM FRACTIONATOR WASTEWATER (STREAMS 155/157)

Organic Compound Class	Concentration, mg/l
Aliphatic Hydrocarbons	290
Ethers	10
Aldehydes and Ketones	25
Nitrites	7.1
Amines	1.6
Thiols, Sulfides, and Disulfides	1.8
Benzene and Substituted Benzene Hydrocarbons	110
Phenols*	280
Fused Polycyclic Hydrocarbons †	64
Fused Non-Alternant Polycyclic Hydrocarbons ‡	38
Heterocyclic Nitrogen Compounds	13
Heterocyclic Oxygen Compounds	180
Heterocyclic Sulfur Compounds (Thiophenes)	11
Total	1,040

Data source: Table 6-30 of Reference 37. These results are preliminary.

* Breakdown for the phenols: 12 mg/l phenol, 13 mg/l o-cresol, 54 mg/l m-cresol, 26 mg/l p-cresol, 75 mg/l C_2 phenol, 24 mg/l C_3 phenol, 15 mg/l methyl naphthol, 19 mg/l C_2 naphthol, 40 mg/l C_3 naphthol, 2 mg/l C_4 naphthol.

† The only fused polycyclic hydrocarbons identified were: naphthalene, methyl naphthalene, phenanthrene, methyl anthracene, methyl phenanthrene, C_2 phenanthrene, and pyrene.

‡ The only fused non-alternant polycyclic hydrocarbons identified were: fluorene, methyl fluorene, C_3 fluorene, C_4 fluorene, and fluoranthene.

3.3.3.1.3 Solid Waste Stream

There is no solid waste stream from liquefaction product fractionation.

3.3.3.2 Solvent Hydrogenation (Product Separation/Purification - Area 3)

In the solvent hydrogenation area, the spent solvent is hydrogenated for reuse; also, liquid products are upgraded through further hydrogenation (see Figure 3-5). The sidestreams from the atmospheric and vacuum fractionators and the liquid distillate from the vacuum fractionator are blended. The blended stream is pumped to process pressure, mixed with hydrogen makeup gas, and fed to the gas-fired solvent hydrogenation feed preheat furnace. The mixture is then passed through a catalytic bed of nickel-molybdenum in the solvent hydrogenation reactor, which operates at an outlet pressure of 11.2 MPa (1605 psig). Recycle quench gas is fed to the reactor in order to absorb the exothermic heat of reaction and control the reactor temperature. The reactor effluent is cooled and sent to the hot separator drum, which separates a vapor and liquid phase at 622°K (660°F) and 11 MPa (1580 psig). The hot separator vapor is cooled, and washed intermittently with water for corrosion control. This mixture is then fed to the cold separator drum, where the hydrogenated solvent, product liquids and water vaporized from the hot separator are condensed and collected, and the unreacted hydrogen and light hydrocarbons formed in the reactor are removed as gases at 316°K (110°F) and 11 MPa (1540 psig).

The vapor from the cold separator drum (Stream 200) is scrubbed with DEA for acid gas removal. A portion of the scrubbed gas is purged to cryogenic hydrogen recovery. The remainder of the scrubbed gas is compressed. Most of the compressed gas is recycled back to the process as quench gas, which is added to the solvent hydrogenation reactor between reactor beds for temperature control. The rest of the compressed gas is heated and recycled as treat gas to the solvent hydrogenation reactors. Ammonia formed in the process is removed with the sour water from the solvent hydrogenation cold separator (Stream 202). Water is formed in the hydrogenation of the spent solvent and liquid products due to the reaction between hydrogen and organically bound oxygen; Stream 202 results from the condensation of this water and water present in the feed to the solvent hydrogenation reactors. The solvent hydrogenation cold separator sour water is sent to wastewater treatment.

The liquid streams from the hot and cold separator drums are sent to a conventional steam stripped solvent fractionator. The products from this solvent stripper include: a sour offgas stream (Stream 251) which consists primarily of hydrogen and methane that will be combined with the atmospheric and Flexicoker fractionator offgas (Streams 151 and 310) and fed to the DEA scrubber for acid gas removal, a naphtha product sent to product recovery, a recycle donor solvent product, and a LSFO product sent to liquid storage. The overhead drum sour water (Stream 252) is sent to wastewater treatment. This sour water results from the condensation of the stripping steam, and is drawn off from the solvent fractionator overhead drum after separation from the overhead gas and the naphtha product. As such, the sour water will contain dissolved gases (H_2S and NH_3) extracted from the overhead gas as well as organics extracted from the overhead gas and the naphtha product.

3.3.3.2.1 Gaseous Waste Stream

There is only one gaseous waste stream from solvent hydrogenation:

- Stream 203 - flue gas from solvent hydrogenation feed preheat furnace.

Flue gas from solvent hydrogenation feed preheat furnace (Stream 203) (Solvent hydrogenation - Area 3)

This flue gas stream is generated from the preheating of the feed to the solvent hydrogenation reactor by the combustion of the low-Btu fuel gas from the Flexicoking unit. Total heat provided by the fuel gas is 0.21 TJ/hr (199.8 MM Btu/hr)(13). The flow rate of the flue gas was calculated to be 3,864 kmol/hr (53,810 SCFM). In Table 3-24, the emission factors used and the estimated emission rates of pollutants for this flue gas stream are presented.

TABLE 3-24. ESTIMATED UNCONTROLLED EMISSIONS FOR FLUE GAS FROM SOLVENT HYDROGENATION FEED PREHEAT FURNACE (STREAM 203)

POLLUTANT	EMISSION FACTOR* ng/J (lb/MM Btu)	EMISSION RATE kg/hr
Particulate matter	2.1 - 6.4 (0.005-0.015)	0.45 - 1.4
NO_x (as NO_2)	52 - 99 (0.12-0.23)	11 - 21
CO	7.3 -10.7 (0.017-0.025)	1.5 - 2.3
Hydrocarbons	1.3 (0.003)	0.27
SO_2	48 (0.11)	10

* Emission factors for particulate matter, NO_x, and hydrocarbons are obtained from AP-42 for natural gas combustion in industrial process heaters (21). By comparison, Exxon estimated emission factors of 3.1 ng/J for particulate matter, and 60 ng/J for NO_x (20). The emission factor for CO is based on 7.3 ng/J from AP-42 and 10.7 ng/J from Exxon. The emission factor for SO_2 is based on 83 ppmv total reduced sulfur in the low-Btu fuel gas (13).

3.3.3.2.2 Liquid Waste Streams

There are two aqueous waste streams from solvent hydrogenation:

- Stream 202 - solvent hydrogenation cold separator wastewater
- Stream 252 - solvent hydrogenation fractionator overhead drum wastewater.

Solvent hydrogenation cold separator wastewater (Stream 202) (Solvent hydrogenation - Area 3)

This wastewater stream is generated from separation of the hydrocarbon liquid, sour water, and vapor in the cold separator drum. The estimated characteristics of the cold separator wastewater are presented in Table 3-25. These estimates were prepared during the Exxon commercial plant design, based on analyses of wastewater samples obtained from several small pilot operating units and computer process synthesis (no data on the species shown in Table 3-25 are available from the EDS pilot plant). In Table 3-26, the concentrations of trace metals in this stream, as determined by Exxon tests of wastewater samples from the 250 ton coal/day pilot plant, are presented. The solvent hydrogenation cold separator wastewater will be combined with other sour water streams before being sent to the sour water stripper/ammonia recovery system for treatment.

Solvent hydrogenation fractionator overhead drum wastewater (Stream 252) (Solvent hydrogenation - Area 3)

The estimated characteristics of this wastewater stream are presented in Table 3-27. Because EPA and Exxon pilot plant test results are available for this stream, these data are presented along with the Exxon design estimates (which were prepared prior to the availability of pilot plant data). Again, there are significant differences between the Exxon design estimates and the pilot plant test results. These differences may be partially explained by the possible differences in process operating conditions and wastewater flow rates. According to Exxon (22), there may be four times more water

TABLE 3-25. ESTIMATED QUALITY OF SOLVENT HYDROGENATION COLD SEPARATOR WASTEWATER (STREAM 202)

Contaminant	Exxon Design Estimate	
	Concentration, ppmw	Flow Rate, kg/hr
H_2S	58,140	730.2
NH_3	54,460	684.0
HCl	0	0
CO_2	0	0
Phenols	1,030	12.9
Organic Acids	0	0
Total flow rate, kg/hr		12,561
Total flow rate, m^3/hr		13.4
Temperature, °K		317
Pressure, MPa		10.7

Data source: Table 1-X-1 of Reference 13 and Attachment 3 of Reference 20.

TABLE 3-26. CONCENTRATIONS OF TRACE METALS IN SOLVENT HYDROGENATION COLD SEPARATOR WASTEWATER (STREAM 202)

Element		Exxon Testing of EDS Pilot Plant, ppmw
Aluminum	Al	0.143
Antimony	Sb	<0.0313
Arsenic	As	(0.0388)
Barium	Ba	0.0412
Beryllium	Be	(0.0017)
Bismuth	Bi	--
Boron	B	31.6*
Cadmium	Cd	<0.0096
Calcium	Ca	2.4
Chromium	Cr	0.147
Cobalt	Co	0.0204
Copper	Cu	0.182
Iron	Fe	1.57
Lead	Pb	(0.0248)
Magnesium	Mg	0.132
Manganese	Mn	0.0644
Mercury	Hg	--
Molybdenum	Mo	<0.0084
Nickel	Ni	0.162
Phosphorus	P	<0.0465
Potassium	K	0.566
Selenium	Se	<0.0212
Silicon	Si	5.84
Silver	Ag	0.0155
Sodium	Na	1.24
Strontium	Sr	0.0041
Tin	Sn	(0.0329)
Titanium	Ti	<0.0011
Tungsten	W	<0.0606
Uranium	U	<0.0293
Vanadium	V	(0.0113)
Zinc	Zn	0.526

Data source: Reference 25.

-- Not analyzed

* Value out of range, needs further dilution

( ) Value may not be significant

TABLE 3-27. CHARACTERISTICS OF SOLVENT HYDROGENATION FRACTIONATOR OVERHEAD DRUM WASTEWATER (STREAM 252) - THREE DATA SOURCES

Contaminant	Exxon Design Estimate	EPA Testing of EDS Pilot Plant	Exxon Testing of EDS Pilot Plant
H_2S	26,720 ppmw	3,905 mg/l	11,692 mg/l
NH_3	18,640 ppmw	Not detected	7,500 mg/l
HCl	0	<1 mg/l	No data
CO_2	0	No data	No data
Phenols	5,160 ppmw	6,852 mg/l	13,000 mg/l
Organic Acids	0	No data	No data
Sulfate	No data	51 mg/l	No data
Cl^-	No data	<1 mg/l	No data
F^-	No data	<1 mg/l	3 mg/l
NO_3^-	No data	2.71 mg/l	No data
NO_2^-	No data	Not detected	No data
SCN^-	No data	210 mg/l	2 mg/l
COD	No data	33,709 mg/l	73,000 mg/l
TDS	No data	166 mg/l	30 mg/l
TSS	No data	7 mg/l	10 mg/l
TOC	No data	4,618 mg/l	16,000 mg/l
Alkalinity (mg/l as $CaCO_3$)	No data	15,681 mg/l	30,500 mg/l
Oil and Grease	No data	3,732 mg/l	No data
Total N	No data	4,040 mg/l	9,100 mg/l
Total S	No data	3,158 mg/l	11,000 mg/l
Flow rate, kg/hr	12,461	No data	No data
Flow rate, m^3/hr	12.7	No data	No data
Temperature, °K	317	No data	No data
Pressure, MPa	0.48	No data	No data

Data source: Table 1-X-1 of Reference 13 and Attachment 3 of Reference 20 for Exxon design estimate, Table 6-45 of Reference 37 for EPA testing results and Reference 23 for Exxon testing results.

condensed (per unit coal input basis) in the pilot plant as in the Exxon design estimate. The Exxon design estimate values will be used in the evaluation of wastewater treatment technology wherever available. As discussed for Stream 152 previously, this is primarily because the two sets of pilot plant data have not yet been integrated into the EDS commercial plant design by Exxon. Exxon and EPA pilot plant test results will only be used to supplement Exxon design estimate values as necessary. In Table 3-28, the characteristics of Stream 252 selected for use in the PCTM water treatment technology evaluation are presented.

The concentrations of trace metals in Stream 252 are presented in Table 3-29. Exxon reported analysis results of individual samples collected from two drums, whereas the EPA analysis results relate to the composite sample from these two drums. Also, the Exxon and EPA samples were taken two weeks apart. Both sets of pilot plant data indicate low levels of trace elements in the wastewater.

In Table 3-30, preliminary results of organic analysis obtained by EPA from the EDS pilot plant wastewater samples are presented. These results indicate that most of the organics present are either phenols or aliphatic hydrocarbons.

The solvent hydrogenation fractionator overhead wastewater will be combined with other sour water streams prior to treatment in the sour water stripper/ammonia recovery system.

3.3.3.2.3 Solid Waste Stream

There is only one solid waste stream from solvent hydrogenation:

- Stream 204 - spent solvent hydrogenation catalyst

Spent solvent hydrogenation catalyst (Stream 204) (Solvent hydrogenation - Area 3)

The discharge rate of spent solvent hydrogenation catalyst is considered proprietary by Exxon. However, in order to obtain an order-of-magnitude

TABLE 3-28. CHARACTERISTICS OF SOLVENT HYDROGENATION FRACTIONATOR OVERHEAD DRUM WASTEWATER USED FOR PCTM WASTEWATER TREATMENT TECHNOLOGY EVALUATION (STREAM 252)

Contaminant	Concentration, ppmw*	Flow Rate kg/hr	Source for Selection of Conc. Value
H_2S	26,720	332.9	Design
NH_3	18,640	232.2	Design
HCl	0	0	Design
CO_2	0	0	Design
Phenols	5,160	64.3	Design
Organic Acids	0	0	Design
Sulfate	52	0.6	EPA Data
F^-	3	0.04	Exxon Data
NO_3^-	2.8	0.03	EPA Data
NO_2^-	Not Detected	0	EPA Data
SCN^-	2	0.02	Exxon Data
COD	54,370	677.5	Avg. Exxon/EPA Data
TDS	31	0.4	Exxon Data
TSS	10	0.1	Exxon Data
TOC	10,510	130.9	Avg. Exxon/EPA Data
Alkalinity (ppmw as $CaCO_3$)	31,080	387.3	Exxon Data
Total flow rate, kg/hr		12,461	
Total flow rate, m^3/hr		12.7	
Temperature, °K		317	
Pressure, MPa		0.48	

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* Note that mg/l values in Table 3-27 have been converted to ppmw values by using the density of the wastewater.

TABLE 3-29. CONCENTRATIONS OF TRACE METALS IN SOLVENT HYDROGENATION FRACTIONATOR OVERHEAD DRUM WASTEWATER (STREAM 252)

Element		EPA Testing of EDS Pilot Plant*	Exxon Testing of EDS Pilot Plant †	
			Drum No. 1	Drum No. 2
Aluminum	Al	0.425	0.316	0.0784
Antimony	Sb	<0.010	<0.0322	(0.0366)
Arsenic	As	0.013	<0.0218	(0.0426)
Barium	Ba	0.0615	0.0385	0.106
Beryllium	Be	<0.003	<0.0012	<0.0012
Bismuth	Bi	<0.50	--	--
Boron	B	1.899	0.909	0.275
Cadmium	Cd	0.025	<0.0099	<0.0099
Calcium	Ca	2.10	1.19	2.21
Chromium	Cr	<0.03	0.0237	0.0207
Cobalt	Co	0.0545	(0.0046)	0.0143
Copper	Cu	<0.015	0.0985	0.0935
Iron	Fe	25.05	0.948	1.39
Lead	Pb	<0.08	<0.0166	<0.0166
Magnesium	Mg	0.205	0.222	0.293
Manganese	Mn	0.225	0.0183	0.0121
Mercury	Hg	Not detected	--	--
Molybdenum	Mo	<0.04	<0.0086	<0.0086
Nickel	Ni	0.057	0.102	0.10
Phosphorus	P	<0.40	<0.0478	<0.0478
Potassium	K	0.345	0.457	0.675
Selenium	Se	<0.001	<0.0218	<0.0218
Silicon	Si	12.8**	7.44	2.59
Silver	Ag	<0.03	<0.0053	<0.0053
Sodium	Na	3.40	1.32	1.98
Strontium	Sr	0.004	0.0037	0.0049
Tin	Sn	<0.03	<0.0166	<0.0166
Titanium	Ti	0.0165	<0.0011	<0.0011
Tungsten	W	--	<0.0624	<0.0624
Uranium	U	--	<0.0302	<0.0302
Vanadium	V	<0.01	(0.0195)	(0.0195)
Zinc	Zn	0.0605	0.264	0.342

Data source: Reference 24 for EPA testing results, Reference 25 for Exxon testing results.

-- Not analyzed

* All EPA values are in units of mg/l. EPA data are for Drums 1 and 2 combined.

† All Exxon values are in units of ppmw

‡ Value out of range, needs further dilution

( ) Value may not be significant.

** EPA silicon values are reported as SiO_2

TABLE 3-30. CONCENTRATIONS OF ORGANIC COMPOUNDS IN SOLVENT HYDROGENATION FRACTIONATOR OVERHEAD DRUM WASTEWATER (STREAM 252)

Organic Compound Class	Concentration, mg/l
Aliphatic Hydrocarbons	1,880
Ethers	200
Aldehydes and Ketones	160
Nitrites	250
Thiols, Sulfides, and Disulfides	6.7
Benzene and Substituted Benzene Hydrocarbons	200
Phenols*	6,910
Heterocyclic Nitrogen Compounds	220
Heterocyclic Oxygen Compounds	35
Heterocyclic Sulfur Compounds (Thiophenes)	280
Total	10,140

Data source: Table 6-42 of Reference 37. These results are preliminary.

* Breakdown for the phenols: 4,600 mg/l phenol, 880 mg/l o-cresol, 1,300 mg/l p-cresol, and 130 mg/l C_2 phenol.

estimate of what quantity of spent catalyst might have to be dealt with, it was assumed that Exxon's estimated storage requirement (6) corresponds to one year of consumption. Based on this assumption, the consumption rate for the solvent hydrogenation catalyst will be 739 Mg/yr for the nickel-molybdenum catalyst and 82 Mg/yr for the inert balls. Thus, approximately 821 Mg of spent solvent hydrogenation catalyst are generated each year. The composition for the spent catalyst is not well defined, but it is expected to contain nickel, molybdenum, carbonaceous material, and sulfides.

3.3.3.3 Gas Treating (Product Separation/Purification - Area 3)

As shown in Figure 3-5, the gas treating area includes a low pressure (L.P.) DEA absorber, a high pressure (H.P.) DEA absorber, and a regenerator for H_2S-rich DEA solution. In the Exxon design, the atmospheric fractionator offgas (Stream 151), solvent stripper offgas (Stream 251), and Flexicoker fractionator offgas (Stream 310) are combined as a low pressure gas stream and treated for H_2S removal in the same L.P. DEA absorber. The treated gas is sent to the cryogenic hydrogen recovery section (Area 4) for recovery of hydrogen, C_1/C_2 light hydrocarbons, and C_3^+ hydrocarbon liquids. The solvent hydrogenation cold separator vapor (Stream 200), a high pressure gas stream, is treated for H_2S removal separately in a H.P. DEA absorber. As described previously in Section 3.3.3.2, a portion of the treated gas is sent to the cryogenic hydrogen recovery section (Area 4), a portion is recycled back to the solvent hydrogenation reactor as quench gas for temperature control, and the remainder is heated and recycled as hydrogen reactant for the solvent hydrogenation reactions.

The H_2S-rich DEA streams from the bottom of the L.P. and H.P. DEA absorbers in Area 3 are combined with the H_2S-rich stream from the bottom of the DEA absorber for the liquefaction cold separator sour gas (Stream 104) in Area 2, and are fed to a common DEA regeneration unit in Area 3. The concentrated H_2S-rich acid gas (Stream 508) from the common DEA regenerator is sent to the sulfur plant. The regenerated H_2S-lean DEA solution is returned to the various individual DEA scrubbers as absorbent.

3.3.3.3.1 Gaseous Waste Stream

There is only one gaseous waste stream from gas treating:

- Stream 508 - acid gas from DEA regenerator.

Acid gas from DEA regenerator (Stream 508) (Gas treating - Area 3)

The composition, flow rates, temperature, and pressure of the acid gas from the common DEA regenerator (Stream 508) for both the base case and the MFS case are presented in Table 3-31. With the exception of the COS composition, all values are based on design estimates provided by Exxon (20). Exxon assumed that all of the COS present in the sour gas streams entering the DEA scrubbers would be hydrolyzed by the DEA, and would thus appear in the DEA regenerator offgas as H_2S. More typically, the less reactive secondary amines such as DEA exhibit COS hydrolysis efficiencies on the order of 70 to 80%, primarily through hydrolysis. Thus, approximately 25% of the COS fed to the DEA scrubbers would be found in the acid gas from the DEA regenerator. The COS concentration in the Stream 508 was estimated by assuming that it is approximately equal to the COS concentration in the acid gas from the DEA unit in the SRC-II commercial plant design (26), or about 510 ppmv. For both the base case and the MFS case, the acid gas from tne DEA regenerator was estimated to contain 64% H_2S.

The characteristics of the sour gas streams entering the DEA scrubbers are presented in Tables 3-32 and 3-33 for the base case and the MFS case, respectively. These values were also based on design estimates provided by Exxon (20). With the exception of Stream 310, the composition and flow rates of the sour gas streams are identical for the base case and MFS case designs.

TABLE 3-31. COMPOSITION AND FLOW RATES OF ACID GAS STREAM FROM EDS COMMERCIAL PLANT PRODUCT SEPARATION AND PURIFICATION OPERATIONS (STREAM 508)

Component	Stream 508 - Acid Gas from DEA Unit	
	Base Case	MFS Case
CO_2, kmol/hr	312.1	296.8
H_2S, kmol/hr	779.4	772.5
NH_3, kmol/hr	47.7	62.2
COS, kmol/hr	0.5	0.5
H_2O, kmol/hr	74.2	73.7
Total, kmol/hr	1,213.9	1,205.7
Temperature, °K	322	322
Pressure, MPa	0.19	0.19

Data source: Attachment 3 of Reference 20 for all components except COS. COS concentration was estimated using SRC-II commercial plant design data (26).

TABLE 3-32. COMPOSITION AND FLOW RATES OF SOUR GAS STREAMS TREATED IN EDS COMMERCIAL PLANT DEA UNIT (ILLINOIS COAL BASE CASE)

Component	Stream 104 Liquefaction Cold Separator Sour Gas	Stream 151 Atmospheric Fractionator Offgas	Stream 200 Solvent Hydrogenation Cold Separator Vapor	Stream 251 Solvent Hydrogenation Fractionator Sour Gas	Stream 310 Flexicoking Fractionator Offgas
H_2, kmol/hr	6,629.6	690.3	17,259.9	670.1	921.4
C_1, kmol/hr	4,576.0	788.5	5,702.5	347.8	644.2
C_2, kmol/hr	871.2	339.1	56.7	8.1	114.5
C_3, kmol/hr	324.9	208.1	8.0	2.0	34.7
C_4, kmol/hr	90.1	100.5	0.5	0.2	10.0
C_5, kmol/hr	23.0	40.4	0.3	0.4	0
C_6^+, kmol/hr	26.9	93.2	23.1	63.6	20.3
N_2, kmol/hr	0	0	0	0	87.7
CO, kmol/hr	60.8	24.2	0	0	80.2
CO_2, kmol/hr	177.8	92.1	0	0	42.2
H_2S, kmol/hr	334.2	370.6	44.0	10.6	21.0
NH_3, kmol/hr	0.6	8.0	34.1	5.4	0
COS, kmol/hr	1.6	0.2	0	0	0
H_2O, kmol/hr	9.2	50.0	19.1	21.6	37.2
Total, kmol/hr	13,125.9	2,805.2	23,148.2	1,129.8	2,013.4
Temperature, °K	317	317	317	317	317
Pressure, MPa	12.7	0.48	10.6	0.48	0.48

Data source: Attachment 3 of Reference 20.

TABLE 3-33. COMPOSITION AND FLOW RATES OF SOUR GAS STREAMS TREATED IN EDS COMMERCIAL PLANT DEA UNIT (ILLINOIS COAL MFS CASE)

Component	Stream 104 Liquefaction Cold Separator Sour Gas	Stream 151 Atmospheric Fractionator Offgas	Stream 200 Solvent Hydrogenation Cold Separator Vapor	Stream 251 Solvent Hydrogenation Fractionator Sour Gas	Stream 310 Flexicoking Fractionator Offgas
H_2, kmol/hr	6,629.6	690.3	17,259.9	670.1	454.1
C_1, kmol/hr	4,576.0	788.5	5,702.5	347.8	317.0
C_2, kmol/hr	871.2	339.1	56.7	8.1	56.5
C_3, kmol/hr	324.9	208.1	8.0	2.0	17.1
C_4, kmol/hr	90.1	100.5	0.5	0.2	5.2
C_5, kmol/hr	23.0	40.4	0.3	0.4	0
C_6+, kmol/hr	26.9	93.2	23.1	63.6	9.4
N_2, kmol/hr	0	0	0	0	45.5
CO, kmol/hr	60.8	24.2	0	0	40.4
CO_2, kmol/hr	177.8	92.1	0	0	26.9
H_2S, kmol/hr	334.2	370.6	44.0	10.6	15.4
NH_3, kmol/hr	0.6	8.0	34.1	5.4	14.7
COS, kmol/hr	1.6	0.2	0	0	0
H_2O, kmol/hr	9.2	50.0	19.1	21.6	18.7
Total, kmol/hr	13,125.9	2,805.2	23,148.2	1,129.8	1,020.9
Temperature, °K	317	317	317	317	317
Pressure, MPa	12.7	0.48	10.6	0.48	0.48

Data source: Attachment 3 of Reference 20.

As shown in Tables 3-32 and 3-33, Stream 104 has a H_2S/CO_2 ratio of 1.9:1 whereas the combined L.P. sour gas has a H_2S/CO_2 ratio of 3.0:1. Carbon dioxide is not present in Stream 200. Enrichment of the H_2S present in these sour gas streams by selective acid gas removal (AGR) processes prior to bulk sulfur recovery, in order to increase the H_2S/CO_2 ratio is therefore not necessary. The data presented in Tables 3-32 and 3-33 can be used to assess the environmental and economic impacts of other acid gas removal systems.

3.3.3.3.2 Liquid Waste Stream

There is no liquid waste stream from gas treating.

3.3.3.3.3 Solid Waste Stream

There is no solid waste stream from gas treating.

3.3.3.4 Product Recovery (Product Separation/Purification - Area 3)

The product recovery operations shown in Figure 3-5 consist of a deethanizer, a debutanizer, and a C_3/C_4 splitter. The compressed gas from cryogenic hydrogen recovery, the cryogenic liquid condensates (C_3^+), and the naphtha distillate from the atmospheric fractionator, the solvent stripper, and the Flexicoker fractionator are fed to the deethanizer. The offgas (C_2^-) from the deethanizer is used as feed to steam reforming for hydrogen generation in the base case, and sent to the cryogenic hydrogen recovery area for C_1/C_2 gas product in the MFS case. The deethanizer bottoms (C_3^+) are fed to the debutanizer where they are fractionated to yield a C_3/C_4 overhead product and a stabilized naphtha ($C_5/400^oF$) bottoms. The naphtha product is cooled and sent to storage. The C_3/C_4 overhead product is treated for removal of trace quantities of hydrogen sulfide and mercaptans, if necessary, by conventional techniques such as caustic scrubbing and water washing. In the current Exxon design, mercaptans present in the spent caustic are oxidized to disulfides, which can then be separated as an oil phase and sent to the Flexicoking unit. The regenerated caustic is reused for C_3/C_4 treating. The treated C_3/C_4 product is fed to the C_3/C_4 splitter. In the C_3/C_4 splitter, the C_3/C_4 stream is fractionated to yield a C_3 LPG overhead and a C_4 LPG bottoms product. The C_4 LPG is cooled and sent to storage. The C_3 LPG is dried and sent to storage.

3.3.3.4.1 Gaseous Waste Stream

There is no gaseous waste stream from product recovery.

3.3.3.4.2 Liquid Waste Stream

There is no liquid waste stream from product recovery.

3.3.3.4.3 Solid Waste Stream

There is no solid waste stream from product recovery.

3.3.4 Liquefaction Residue Processing/Hydrogen Production Area (Area 4)

Vacuum-bottoms slurry from the product separation and purification area (Area 3) is utilized to generate additional liquid products (e.g., naphtha and low sulfur fuel oil), fuel gas and either light gases or syngas (by partial oxidation) for hydrogen production. In the MFS case, high Btu gas is also produced for sale. The principal operations involved in processing vacuum-bottoms are Flexicoking* and hydrogen generation. These operations are presented schematically in Figures 3-6 and 3-7 for the base and MFS cases, respectively, and are described in the ensuing sections.

3.3.4.1 Flexicoking (Liquefaction Residue Processing/Hydrogen Production - Area 4)

The Flexicoking unit converts vacuum-bottoms slurry through reactions with steam and air into additional liquid and gas products, and generates low Btu fuel gas for process consumption. Flexicoking is a low pressure process (<0.45 MPa; <65 psia) consisting of a fluid-bed reactor (coker), a heater, and a gasifier. Vacuum-bottoms slurry is coked in the fluid-bed reactor section in the presence of steam at temperatures of 755-920°K (900-1200°F) to yield liquid and gas products and coke. Reaction heat is supplied by a circulating stream of coke which transports heat from the heater to the reactor vessel. Temperatures in the heater are maintained by circulating gas and solids from the gasifier. Coke produced in the Flexicoking reactor circulates to the heater where it is devolatilized to yield a light hydrocarbon gas and residual coke. The residual coke is then fed to the gasifier, where it is reacted with air and steam at 1100-1250°K (1500-1800°F) to form low Btu fuel gas.

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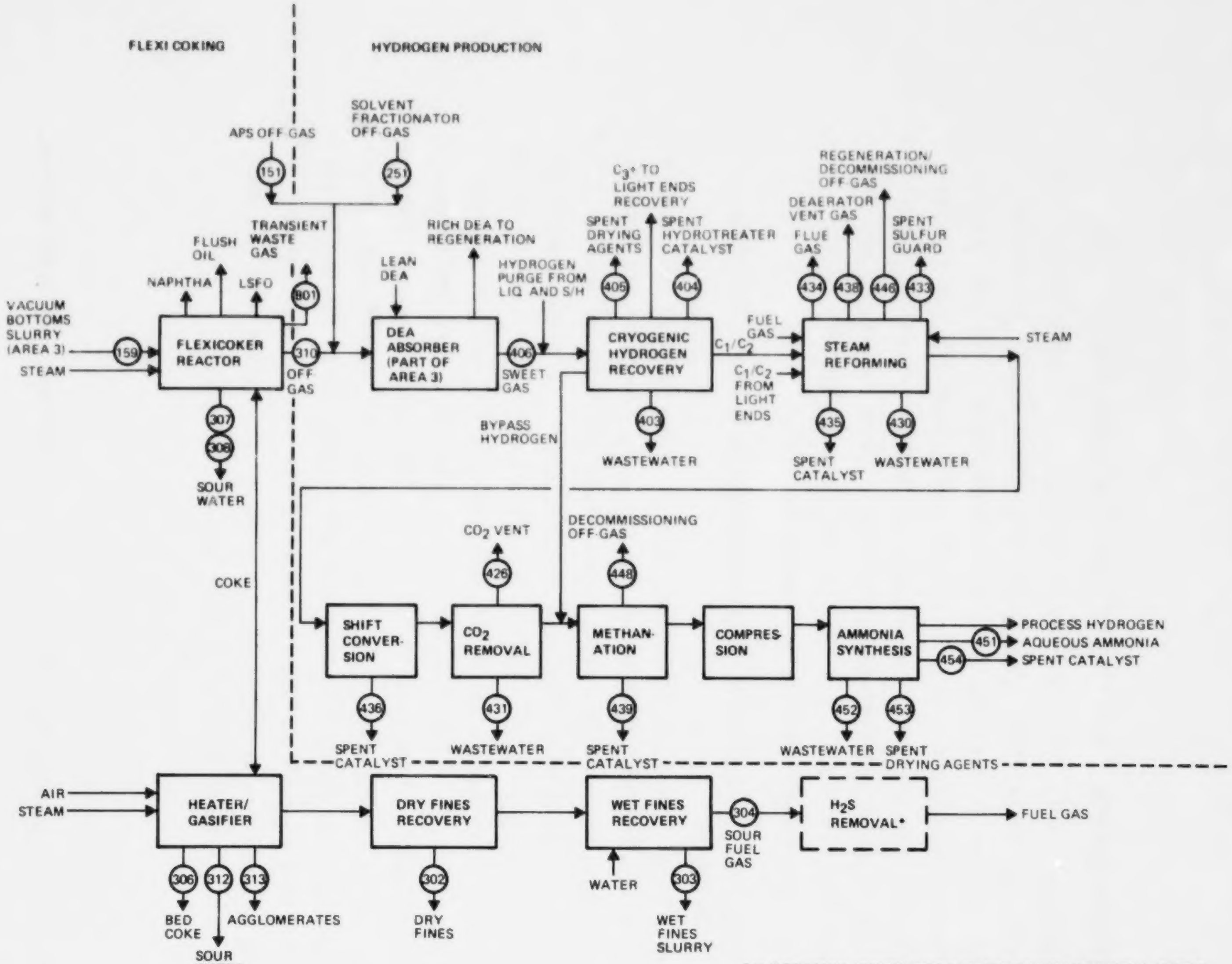
* Flexicoking is the service mark for a proprietary process developed by Exxon.


FLEXI COKING
HYDROGEN PRODUCTION
APS OFF GAS
151
SOLVENT FRACTIONATOR OFF GAS
251
VACUUM BOTTOMS SLURRY (AREA 3)
159
STEAM
FLEXICOKER REACTOR
NAPHTHA
FLUSH OIL
LSFO
TRANSIENT WASTE GAS
901
310
OFF GAS
307
308
SOUR WATER
COKE
LEAN DEA
DEA ABSORBER (PART OF AREA 3)
RICH DEA TO REGENERATION
406
SWEET GAS
HYDROGEN PURGE FROM LIQ AND S/H
BYPASS HYDROGEN
SPENT DRYING AGENTS
405
C_3+ TO LIGHT ENDS RECOVERY
SPENT HYDROTREATER CATALYST
404
CRYOGENIC HYDROGEN RECOVERY
403
WASTEWATER
C_1/C_2
FUEL GAS
C_1/C_2 FROM LIGHT ENDS
FLUE GAS
434
DEAERATOR VENT GAS
438
REGENERATION/ DECOMMISSIONING OFF-GAS
446
SPENT SULFUR GUARD
433
STEAM REFORMING
STEAM
435
SPENT CATALYST
430
WASTEWATER
SHIFT CONVERSION
436
SPENT CATALYST
CO_2 VENT
426
CO_2 REMOVAL
431
WASTEWATER
DECOMMISSIONING OFF-GAS
448
METHANATION
439
SPENT CATALYST
COMPRESSION
AMMONIA SYNTHESIS
PROCESS HYDROGEN
451
AQUEOUS AMMONIA
454
SPENT CATALYST
452
WASTEWATER
453
SPENT DRYING AGENTS
AIR
STEAM
HEATER/ GASIFIER
306
BED COKE
312
SOUR WATER
313
AGGLOMERATES
DRY FINES RECOVERY
302
DRY FINES
WATER
WET FINES RECOVERY
303
WET FINES SLURRY
304
SOUR FUEL GAS
H_2S REMOVAL*
FUEL GAS

* H_2S REMOVAL REPRESENTS A SULFUR EMISSION CONTROL OPTION

Figure 3-6. Block flow diagram for EDS liquefaction residue processing and hydrogen production (Area 4) - base case

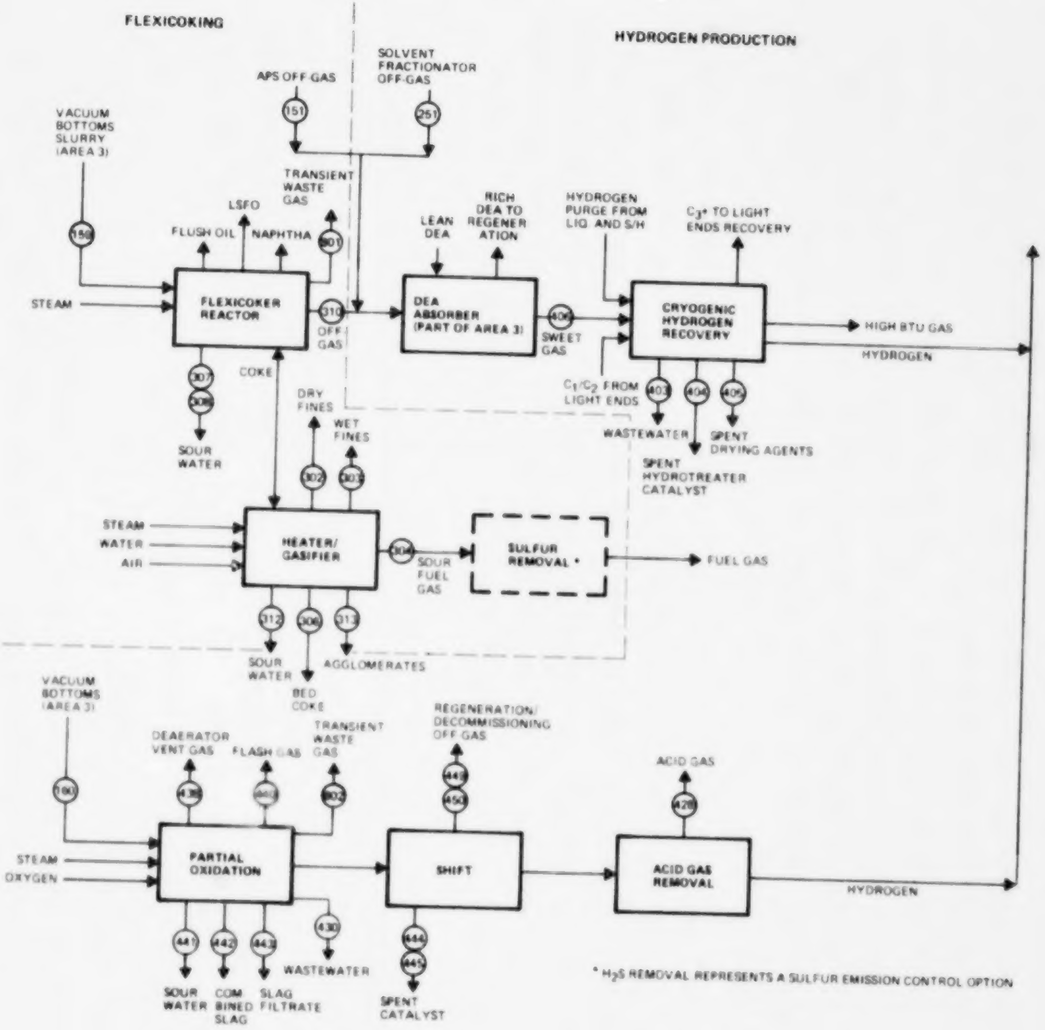

FLEXICOKING
HYDROGEN PRODUCTION
VACUUM BOTTOMS SLURRY (AREA 3)
STEAM
FLUSH OIL
LSFO
NAPHTHA
TRANSIENT WASTE GAS
801
APS OFF-GAS
151
SOLVENT FRACTIONATOR OFF-GAS
251
FLEXICOKER REACTOR
310
OFF GAS
307
308
SOUR WATER
COKE
LEAN DEA
RICH DEA TO REGENERATION
DEA ABSORBER (PART OF AREA 3)
406
SWEET GAS
HYDROGEN PURGE FROM LIQ. AND S/H
C_1/C_2 FROM LIGHT ENDS
C_3^+ TO LIGHT ENDS RECOVERY
CRYOGENIC HYDROGEN RECOVERY
HIGH BTU GAS
HYDROGEN
403
404
405
WASTEWATER
SPENT HYDROTREATER CATALYST
SPENT DRYING AGENTS
DRY FINES
WET FINES
302
303
STEAM
WATER
AIR
HEATER/ GASIFIER
304
SOUR FUEL GAS
SULFUR REMOVAL *
FUEL GAS
312
306
313
SOUR WATER
BED COKE
AGGLOMERATES
VACUUM BOTTOMS (AREA 3)
STEAM
OXYGEN
DEAERATOR VENT GAS
FLASH GAS
TRANSIENT WASTE GAS
438
802
PARTIAL OXIDATION
441
442
443
430
SOUR WATER
COMBINED SLAG
SLAG FILTRATE
WASTEWATER
REGENERATION/ DECOMMISSIONING OFF GAS
449
450
SHIFT
444
445
SPENT CATALYST
ACID GAS
428
ACID GAS REMOVAL
HYDROGEN
* H_2S REMOVAL REPRESENTS A SULFUR EMISSION CONTROL OPTION

Figure 3-7. Block flow diagram for EDS liquefaction residue processing and hydrogen production (Area 4) - MFS case

Vapor products from the Flexicoking reactor are cooled and separated, and heavy organics and solids are recycled to the reactor. The lighter organics are fractionated to separate naphtha from low sulfur fuel oil and wash oil. Olefinic coker gas from fractionation is cooled, compressed, and sent to hydrogen recovery. Coker gas (Stream 310) consists primarily of light hydrocarbons, hydrogen and carbon oxides with lesser amounts of nitrogen, hydrogen sulfide and ammonia. The composition of the coker gas has been presented in Tables 3-32 and 3-33.

Raw fuel gas from the heater/gasifier system is treated for particulate removal. In the Exxon design, entrained particulate consisting of ash and approximately 20% residual coke is removed by sequential dry and wet (venturi scrubber) removal systems. Wet fines (Stream 303) are recovered as a 6% solids slurry and dewatered to 40% solids prior to disposal. The sour fuel gas stream (Stream 304) may be further treated for sulfur removal prior to use as plant fuel.

Ash and residual coke (Stream 306) are removed on a continuous basis from the heater/gasifier system at an ash-to-coke weight ratio of approximately two. Chunk coke or agglomerates (Stream 313) are removed from the system for disposal on a daily basis. There might also be air emissions associated with the recovery of coke and coke fines from the Flexicoking operations, but the rates of these emissions cannot be readily determined.

3.3.4.1.1 Gaseous Waste Streams

There are two gaseous waste streams from Flexicoking:

- Stream 304 - Flexicoking gasifier/heater sour fuel gas
- Stream 801 - transient waste gas from Flexicoking.

Flexicoking gasifier/heater sour fuel gas (Stream 304) (Flexicoking - Area 4)

This gaseous stream is generated in the gasifier vessel where coke formed in the Flexicoking reactor is gasified with steam and air. The product sour fuel gas, after passing through the heater and waste heat boilers, contains

0.4-0.5% sulfur as H_2S and COS, as a result of the residual sulfur that was present in the coke. With a heating value of 4.7 MJ/Nm^3 (125 Btu/SCF, dry basis), combustion of this sour fuel gas would result in SO_2 emissions of approximately 2550 ng/J (5.9 lb/MM Btu). Hence, removal of reduced sulfur compounds from the sour fuel gas prior to combustion in process heaters and boilers is a pollution control option that may be considered. Characterization data for the sour fuel gas stream under the base case and MFS case designs are presented in Table 3-34. These characteristics are based on EDS design information provided by Exxon (20). The sour fuel gas is a large stream with a low H_2S/CO_2 ratio (<0.04) and cannot be conveniently treated in DEA adsorbers because the resultant acid gas would not be sufficiently concentrated as feed to Claus sulfur recovery plants.

Transient waste gas from Flexicoking (Stream 801) (Flexicoking - Area 4)

Transient waste gas is generated from Flexicoking operations during startup, shutdown, and upset conditions. For the three parallel Flexicoking units in the base case design, it is assumed that transient conditions generating waste gases would occur 30 times per year per unit due to unscheduled outages (13), and up to 3 hours in duration each time. Also, outage for the Flexicoking units will not occur simultaneously. Similar assumptions are made for the two Flexicoking units in the MFS case design. The transient waste gas from the Flexicoking unit is anticipated to be similar in composition and flow rate to the Flexicoker fractionator off-gas (Stream 310). Characterization data for this transient waste gas stream are presented in Table 3-35. The estimated flow rate of this stream during periods of release is 671 kmol/hr for the base case design and 511 kmol/hr for the MFS case design.

3.3.4.1.2 Liquid Waste Streams

There are three aqueous waste streams from Flexicoking:

- Stream 307 - Flexicoking recontacting drum wastewater

TABLE 3-34. ESTIMATED COMPOSITION AND FLOW RATES OF SOUR FUEL GAS FROM THE FLEXICOKER HEATER/GASIFIER UNIT (STREAM 304)

Component	Stream 304 - Sour Fuel Gas	
	Base Case	MFS Case
H_2, kmol/hr	9,821.7	5,364.3
C_1, kmol/hr	831.3	416.2
N_2, kmol/hr	25,757.1	12,471.0
CO, kmol/hr	9,834.9	4,735.5
CO_2, kmol/hr	6,343.0	3,141.0
H_2S, kmol/hr	229.2	121.0
COS, kmol/hr	5.9	2.5
H_2O, kmol/hr	1,958.2	973.3
Total, kmol/hr	54,781.3	27,224.8
Temperature, °K	316	316
Pressure, MPa	0.2	0.2

Data source: Characterization data are based upon EDS design information from Attachments 3 and 4 of Reference 20.

TABLE 3-35. ESTIMATED CHARACTERISTICS OF TRANSIENT WASTE GAS FROM FLEXICOKING UNITS (STREAM 801)

Component	Base Case		MFS Case	
	Volume Percent	Flow Rate, kmol/hr	Volume Percent	Flow Rate, kmol/hr
H_2	45.8	307.1	44.5	227.1
C_1	32.0	214.7	31.1	158.5
C_2	5.7	38.2	5.5	28.3
C_3	1.7	11.6	1.7	8.6
C_4	0.5	3.3	0.5	2.6
C_5	0	0	0	0
C_6^+	1.0	6.8	0.9	4.7
N_2	4.4	29.2	4.5	22.8
CO	4.0	26.7	4.0	20.2
CO_2	2.1	14.1	2.6	13.5
H_2S	1.0	7.0	1.5	7.7
NH_3	0	0	1.4	7.4
H_2O	1.8	12.4	1.8	9.4
Total flow rate		671.1		510.8
Temperature, °K		317		317
Pressure, MPa		0.48		0.48

Data source: Characteristics of Stream 310 in Attachments 3 and 4 in Reference 20.

Estimated period of release: 30 occurrences per year for each Flexicoking unit; up to 3 hours duration per occurrence.

- Stream 308 - Flexicoking fractionator overhead drum wastewater
- Stream 312 - Flexicoking heater overhead drum wastewater.

Flexicoking recontacting drum wastewater (Stream 307) (Flexicoking - Area 4)

This wastewater stream is generated from compression of the Flexicoking fractionator off-gas leaving the overhead drum to 0.48 MPa (55 psig), followed by cooling to 317°K (110°F), partial condensation, and liquid phase separation in the recontacting drum. Thus, Stream 307 is the additional water vapor condensed from the Flexicoking fractionator off-gas (at a higher pressure), after an initial amount of water vapor is condensed and removed as Stream 308 (at a lower pressure). Streams 307 and 308 both result from unreacted steam added to the Flexicoking reactor. Streams 307 and 308 will contain dissolved acid gases, ammonia, and organics due to intimate contact with the Flexicoking fractionator off-gas. The hydrogen sulfide and ammonia result from inorganically bound sulfur and nitrogen in coal, which remained in vacuum bottoms and were released during Flexicoking. Compared with Stream 308, Stream 307 is relatively low in organics because most soluble organics have been removed with Stream 308. The estimated characteristics of this wastewater stream are presented in Table 3-36. These estimated characteristics are based on Exxon's design information (13,20). The composition of this wastewater stream is the same for the base case and MFS case designs. However, only half of the vacuum bottoms slurry is sent to the Flexicoking unit in the MFS case design. Therefore, the flow rate in the MFS case design is only half that of the base case design. Stream 307 will likely be combined with other sour water streams before being sent to wastewater treatment.

Flexicoking fractionator overhead drum wastewater (Stream 308) (Flexicoking - Area 4)

This wastewater stream is generated from initial condensation of water vapor and heavy organics present in the Flexicoking fractionator off-gas at 0.18 MPa (11 psig) and 317°K (110°F), followed by phase separation in the overhead drum. As discussed for Stream 307 previously, Stream 308 results from

TABLE 3-36. ESTIMATED QUALITY OF FLEXICOKING RECONTACTING DRUM WASTEWATER (STREAM 307)

Contaminant	Base Case		MFS Case	
	Concentration ppmw	Flow Rate, kg/hr	Concentration ppmw	Flow Rate, kg/hr
H_2S	17,640	23.1	17,640	11.6
NH_3	23,860	31.3	23,860	15.6
HCl	0	0	0	0
CO_2	25,930	34.0	25,930	17.0
Phenols	16	9.5	16	4.8
Organic Acids	0	0	0	0
Flow rate, kg/hr		1,312		656
Flow rate, m^3/hr		1.4		0.67
Temperature, °K		316		316
Pressure, MPa		0.5		0.5

Data source: Table 1-X-1 of Reference 13, Attachments 3 and 4 of Reference 20.

condensation of unreacted steam that was added to the Flexicoking reactor. The hydrogen sulfide and ammonia present in Stream 308 result from inorganically bound sulfur and nitrogen (which had remained in the vacuum bottoms) released in the Flexicoking reactor. Stream 308 is higher in organics than Stream 307 because most of the heavier organics in the Flexicoker fractionator overhead are condensed at the lower pressure and removed with Stream 308 as water-soluble phenols. The estimated characteristics of this wastewater stream, based on Exxon design information, are presented in Table 3-37. Again, the flow rate of this wastewater stream for the MFS design is half that for the base case design, although the stream compositions are identical for the two designs. Stream 308 will likely be combined with other sour water streams before being sent to wastewater treatment.

Flexicoking gasifier/heater overhead wastewater (Stream 312) (Flexicoking - Area 4)

This wastewater stream results from condensation of the water vapor (unreacted steam fed to the gasifier) present in the sour fuel gas stream from the Flexicoking gasifier/heater. As expected, the wastewater stream is fairly well saturated with H_2S, NH_3 and CO_2 as shown in the estimated characteristics presented in Table 3-38, due to the presence of these species in the sour fuel gas. Organic levels in this waste stream are low because the sour fuel gas contains mostly lighter hydrocarbon gases. Similar to other wastewater streams from Flexicoking, the flow rate of Stream 312 for the MFS design is half that for the base case design. Stream compositions for the two designs are again identical. Stream 312 will also be combined with other sour water streams prior to wastewater treatment.

3.3.4.1.3 Solid Waste Streams

There are four solid waste streams from Flexicoking:

- Stream 302 - Flexicoking gasifier/heater dry fines
- Stream 303 - Flexicoking gasifier/heater wet fines

TABLE 3-37. ESTIMATED QUALITY OF FLEXICOKING FRACTIONATOR OVERHEAD DRUM WASTEWATER (STREAM 308)

Contaminant	Base Case		MFS Case	
	Concentration, ppmw	Flow Rate, kg/hr	Concentration ppmw	Flow Rate, kg/hr
H_2S	1,770	176.9	1,770	88.5
NH_3	2,140	213.6	2,140	106.8
HCl	145	14.5	145	7.3
CO_2	2,620	261.3	2,620	130.6
Phenols	3,890	388.7	3,890	194.4
Organic Acids	0	0	0	0
Flow rate, kg/hr		99,910		49,955
Flow rate, m^3/hr		100.9		50.4
Temperature, °K		316		316
Pressure, MPa		0.9		0.9

Data source: Table 1-X-1 of Reference 13, Attachments 3 and 4 of Reference 20.

TABLE 3-38. ESTIMATED QUALITY OF FLEXICOKING HEATER OVERHEAD DRUM WASTEWATER (STREAM 312)

Contaminant	Base Case		MFS Case	
	Concentration ppmw	Flow Rate, kg/hr	Concentration ppmw	Flow Rate, kg/hr
H_2S	6,020	171.0	6,020	85.5
NH_3	13,930	395.5	13,930	197.8
HCl	0	0	0	0
CO_2	21,680	615.5	21,680	307.8
Phenols	16	0.5	16	0.2
Organic Acids	0	0	0	0
Flow rate, kg/hr		28,400		14,200
Flow rate, m^3/hr		28.7		14.3
Temperature, °K		316		316
Pressure, MPa		0.7		0.7

Data source: Table 1-X-1 of Reference 13, Attachments 3 and 4 of Reference 20.

- Stream 306 - Flexicoking gasifier/heater bed coke
- Stream 313 - Flexicoking heater/reactor chunks/agglomerates.

Essentially all of the ash in the coal fed to the EDS commercial plant will appear in these four streams for the base case design. In the MFS case, about half of the ash will appear in these Flexicoking solid waste streams, and the other half will appear as slag from the partial oxidation unit in the hydrogen production area.

Flexicoking gasifier/heater dry fines (Stream 302) (Flexicoking - Area 4)

This solid waste stream results from the collection of particulate matter from the Flexicoking heater overhead in tertiary cyclones.* The heater overhead consists of a light hydrocarbon gas from the devolatilization of coke in the heater, as well as low-Btu gas generated in the gasifier and passed upward through the heater. Based on Exxon's estimate, the tertiary cyclones dry fines are generated continuously at the rate of 137,950 Mg/yr for the base case design and 65,670 Mg/yr for the MFS case design (13). These dry fines have a bulk density of 0.61 Mg/m^3 (38.0 lb/ft^3), a particle density of 1.94 Mg/m^3 (121.0 lb/ft^3), and contain approximately 80 wt % coal ash and 20 wt % carbonaceous material (6,13). In Table 3-39, the results of leaching tests on Flexicoking dry fines using the 12/18/1978 EPA Extraction Procedure (EP) are presented (20). The analysis results show that concentrations of all elements in the leachate are below limits set by RCRA criteria. RCRA testing will need to be conducted on dry fines produced in commercial EDS facilities to confirm these results.

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* These particulate controls are considered an integral part of the EDS process, primarily because particulates need to be removed from the sour fuel gas before it can be used as fuel in process heaters.

TABLE 3-39. ELEMENTAL ANALYSES OF LEACHATES DERIVED FROM EPA DRAFT EXTRACTION PROCEDURE APPLIED TO EDS FLEXICOKING GASIFIER/HEATER DRY FINES* (STREAM 302)

Element	Concentration in Leachate, ppmw	RCRA Criteria[†] ppmw
F^-	0.3	140-240
NO_3^-	<3	4430
Ag	N.D. (<0.002)[‡]	5.0
Al	17.5	
As	N.D. (<0.055)	5.0
B	12.1	
Ba	1.76	100
Be	0.003	
Ca	2.060	
Cd	0.0425	1.0
Co	0.155	
Cr	0.0042	5.0
Cu	0.44	
Fe	3.54	
Hg	0.0004	0.2
K	3.89	
Mg	328	
Mn	2.3	
Mo	N.D. (<0.009)	
Na	22	
Nb	N.D. (<0.016)	
Ni	1.19	
P	0.3	
Pb	[0.0462]**	5.0
Pt	[0.0364]	
Sb	No test	
Se	N.D. (<0.036)	1.0
Si	25.5	
Sn[††]	0.889	
Ti	0.0252	
Tl	N.D. (<0.012)	
U	N.D. (<0.051)	
V	N.D. (<0.0098)	
Zn	1.68	

Data source: Attachment 10 of Reference 20.

[*] Extraction procedure based upon draft EPA procedures proposed September 12, 1978 and December 18, 1978. Results based upon solid wastes obtained from runs on Illinois coal EDS vacuum bottoms in small Flexicoking units.

[†] RCRA criteria are 100 times the National Interim Primary Drinking Water Standards.

[‡] N.D. indicates that the element was not detected. Detection limits are provided in parentheses.

[**] Bracketed concentrations may not be analytically significant.

[††] Values for tin may be too high due to the method of sample storage.

Flexicoking gasifier/heater wet fines (Stream 303) (Flexicoking - Area 4)

This solid waste stream results in the Exxon design from the collection of additional particulate matter from the Flexicoking heater overhead downstream of the tertiary cyclones. The particulate matter is collected in a water slurry using a venturi scrubber,* and the slurry is subsequently dewatered to generate a waste containing 40 wt % solids. Based on Exxon's estimate, these wet fines are generated continuously at the rate of 344,870 Mg/yr (40 wt % solids basis) for the base case design, and 164,160 Mg/yr (40 wt % solids basis) for the MFS case design (13). The wet fines have the same particle density as the dry fines (1.94 Mg/m^3), and also contain approximately 80 wt % coal ash and 20 wt % carbonaceous material (6,13). The leaching characteristics of the wet fines are expected to be similar to the leaching characteristics of the dry fines. However, RCRA testing will need to be conducted on wet fines produced in commercial EDS facilities to confirm this.

Flexicoking gasifier/heater bed coke (Stream 306) (Flexicoking - Area 4)

This is a waste stream that results from continuous purge of the recirculating solids which circulate between the reactor and the heater and between the gasifier and the heater. The purge is necessary to avoid buildup of metals and ash in the system. The generation rate of the bed coke purge is estimated to be 707,100 Mg/yr for the base case design and 363,030 Mg/yr for the MFS case design (13). The bed coke purge contains approximately 61 wt % coal ash and 39 wt % carbonaceous material (13). The results of leaching tests on this waste stream using the 12/18/1978 EPA Extraction Procedure (EP) are presented in Table

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* These particulate controls are considered an integral part of the EDS process, primarily because particulates need to be removed from the sour fuel gas before it can be used as fuel in process heaters.

3-40. These results show that the concentration levels of all elements in the leachate from the bed coke purge are very low and well below RCRA criterial limits.

Flexicoking heater/reactor chunks/agglomerates (Stream 313) (Flexicoking - Area 4)

This is an intermittent waste stream generated daily from purge of coke chunks/agglomerates from the Flexicoking heater and reactor to avoid buildup of oversize coke. On annual basis, the generation rate of this solid waste is estimated to be 19,860 Mg for the base case design and 9,660 Mg for the MFS case design (13). The chunks/agglomerates have a bulk density of 1 Mg/m^3 (65.0 lb/ft^3), and should have composition similar to the bed coke purge (Stream 306). Particle size distribution for the chunks/agglomerates was estimated by Exxon to be as follows: 50 wt % between 28 mesh and 1/4 in.; 40 wt % between 1/4 in. and 1 in.; and 10 wt % greater than 1 in. (6). Leaching characteristics of the chunks/agglomerates using the EPA Extraction Procedure (EP) are presented in Table 3-41. As with other Flexicoking solid wastes, the concentration levels of the pertinent trace elements, measured on chunks/agglomerates from experimental Flexicoking runs, are extremely low and well below RCRA EP criteria. RCRA testing will need to be conducted on chunks/agglomerates produced in commercial EDS facilities in order to confirm these results.

3.3.4.2 Hydrogen Production (Liquefaction Residue Processing/Hydrogen Production - Area 4)

A major difference between the base and MFS cases is the method of hydrogen production. Both cases obtain a portion of the make-up hydrogen requirement by cryogenic separation of hydrogen from hydrogen-rich gas streams leaving the liquefaction, solvent hydrogenation, product distillation, solvent fractionation and Flexicoking areas. However, the balance of the hydrogen requirement is generated by steam reforming of light hydrocarbon gases in the base case and partial oxidation of vacuum-bottoms slurry in the MFS case.

TABLE 3-40. ELEMENTAL ANALYSES OF LEACHATES DERIVED FROM EPA DRAFT EXTRACTION PROCEDURE APPLIED TO EDS FLEXICOKING GASIFIER/HEATER BED COKE* (STREAM 306)

Element	Concentration in Leachate, ppmw	RCRA Criteria[†] ppmw
F^-	0.94	140-240
NO_3	<3	4430
Ag	N.D. (<0.0026)[‡]	5.0
Al	0.112	
As	N.D. (<0.036)	5.0
B	3.89	
Ba	0.032	100
Be	N.D. (0.0009)	
Ca	38.9	
Cd	N.D. (<0.0066)	1.0
Co	N.D. (<0.0061)	
Cr	N.D. (<0.003)	5.0
Cu	0.47	
Fe	0.0215	
Hg	0.0009	0.2
K	5.41	
Mg	1.04	
Mn	0.0587	
Mo	N.D. (<0.019)	
Na	3.7	
Nb	N.D. (<0.012)	
Ni	N.D. (<0.01)	
P	N.D. (<0.051)	
Pb	N.D. (<0.019)	5.0
Pt	N.D. (<0.027)	
Sb	0.0443	
Se	N.D. (<0.055)	1.0
Si	2.89	
Sn[††]	N.D. (<0.024)	
Ti	0.0095	
Tl	N.D. (<0.027)	
U	N.D. (<0.063)	
V	0.0181	
Zn	N.D. (<0.0035)	

Data source: Attachment 10 of Reference 20.

[*] Extraction procedure based upon draft EPA procedures proposed September 12, 1978 and December 18, 1978. **Results based upon solid wastes obtained from runs on Illinois coal EDS vacuum bottoms in small Flexicoking units.**

[†] RCRA criteria are 100 times the **National Interim Primary Drinking Water Standards.**

[‡] N.D. indicates that the element was not detected. Detection limits are provided in parentheses.

[††] Values for tin may be too high due to the method of sample storage.

TABLE 3-41. ELEMENTAL ANALYSES OF LEACHATES DERIVED FROM EPA DRAFT EXTRACTION PROCEDURE APPLIED TO EDS FLEXICOKING HEATER/REACTOR CHUNKS/AGGLOMERATES* (STREAM 313)

Element	Concentration in Leachate, ppmw	RCRA Criteria,† ppmw
F^-	0.17	140-240
NO_3	<3	4430
Ag	N.D. (<0.002) ‡	5.0
Al	0.485	
As	N.D. (<0.055)	5.0
B	0.765	
Ba	0.0407	100
Be	N.D. (<0.0004)	
Ca	25.4	
Cd	N.D. (<0.0048)	1.0
Co	0.658	
Cr	0.0019	5.0
Cu	N.D. (<0.0006)	
Fe	2.19	
Hg	N.D. (<0.0002)	0.2
K	1.55	
Mg	0.403	
Mn	0.0496	
Mo	N.D. (<0.009)	
Na	44	
Nb	N.D. (<0.016)	
Ni	0.236	
P	[0.0941]**	
Pb	N.D. (<0.023)	5.0
Pt	N.D. (<0.021)	
Sb	N.D. (<0.076)	
Se	N.D. (<0.036)	1.0
Si	0.979	
Sn	N.D. (<0.013)	
Ti	N.D. (<0.001)	
Tl	N.D. (<0.012)	
U	N.D. (<0.051)	
V	N.D. (<0.0098)	
Zn	N.D. (<0.0301)	

Data source: Attachment 10 of Reference 20.

* Extraction procedure based upon draft EPA procedures proposed September 12, 1978 and December 18, 1978. **Results based upon solid wastes obtained from runs on Illinois coal EDS vacuum bottoms in small Flexicoking units.**

† RCRA criteria are 100 times the **National Interim Primary Drinking Water Standards.**

‡ N.D. indicates that the element was not detected. Detection limits are provided in parentheses.

** Bracketed concentrations may not be analytically significant.

†† Values for tin may be too high due to the method of sample storage.

In the Exxon design for the EDS base case, shown in Figure 3-6, hydrogen production consists of: a hydrogen recovery section, a hydrogen generation section, and an ammonia synthesis section. The hydrogen recovery section cryogenically separates hydrogen, C_1-C_3 steam reformer feed, and C_3^+ product from the high-pressure purge gases and low-pressure off-gases leaving the liquefaction, solvent hydrogenation, production distillation, and Flexicoking areas. The low-pressure off-gases, which contain significant quantities of H_2S and CO_2, are treated by low pressure DEA scrubbing in Area 3 prior to cryogenic separation. After DEA scrubbing, the low-pressure gases are compressed from 0.45 MPa (50 psig) to 5.2 MPa (740 psig), and hydrotreated over nickel-molybdenum catalyst to remove olefins present in the coker off-gas. Spent hydrotreater catalyst (Stream 404) is removed periodically as a result of deactivation by fouling (e.g., carbon deposition) or poisoning (e.g., chemisorption of residual hydrogen sulfide present). The off-gases are then cooled to 297°K (75°F) which results in the condensation of most of the water vapor present in knockout drums (Stream 403). The high-pressure purge gases which come from the high-pressure DEA scrubber in Area 3 are depressurized and cooled to 297°K (75°F), and then combined with the cleaned, low-pressure off-gases. The combined gas stream is dried with zeolites to remove the remaining water, H_2S and CO_2 prior to being sent to the cold box, where the cryogenic separation takes place. Spent zeolites (Stream 405) are removed periodically as they become saturated with water, H_2S and CO_2.

Hydrogen generation in the EDS base case design involves steam reforming, shift conversion, carbon dioxide removal and removal of residual carbon oxides. Feed streams to the reformer are the C_3^- hydrocarbons after hydrogen removal in cryogenic hydrogen recovery, and the C_1 and C_2 hydrocarbons from light ends recovery. These streams are passed through a ZnO sulfur guard for trace sulfur removal prior to steam reforming. Reforming of the hydrocarbon feed is performed over a nickel-urania catalyst at approximately 2 MPa (300 psia) and

1090°K (1500°F), reacting the steam and the light hydrocarbons to yield hydrogen and carbon oxides (CO and CO_2). Reaction heat is provided by burning low sulfur fuel gas; waste heat is recovered in steam production for the reformer feed and for other process uses. Additional hydrogen is produced by high temperature (>622°K; >660°F) shift conversion (of CO + H_2O to H_2 and CO_2) over an iron oxide catalyst.

The gas following the reforming and shift reactions consists of hydrogen with high levels of CO_2 and lower levels of CO, water vapor, and light hydrocarbons. Bulk removal of carbon dioxide from the shift gas can be accomplished by an acid gas removal (AGR) process; the Catacarb process is employed in the current Exxon design. This process is a hot potassium carbonate process utilizing amine borates to increase the carbonate solution activity. A Catacarb or alternative AGR unit will remove most of the carbon dioxide and water vapor along with a small amount of hydrogen, carbon monoxide and methane. Any ammonia which may be generated by reaction of hydrogen and nitrogen during hydrogen generation will also be removed. Hence, regeneration of rich Catacarb (or alternative AGR) solution will produce a carbon dioxide rich off-gas containing water vapor and small quantities of hydrogen, carbon monoxide, methane and possibly ammonia.

Hydrogen rich gas from the base case carbon dioxide removal unit is combined with the hydrogen from cryogenic separation prior to removal of trace carbon oxides. Methanation of the combined hydrogen stream is employed to remove trace carbon oxides by converting them to methane, through reaction with H_2 and typically over a nickel oxide catalyst at 588 to 638°K (599 to 689°F) and 3 MPa (420 psig) (33). In the methanation step, carbon oxide levels are reduced to 10 ppmv or less to prevent catalyst poisoning in the subsequent nitrogen removal/ammonia synthesis step. Methanation is followed by cooling and compression, steps in which most of the water vapor is condensed and removed in knockout drums (part of Stream 430).

In the EDS base design, a proprietary ammonia synthesis step is employed to remove nitrogen present in the hydrogen-rich gas from methanation. This nitrogen is introduced into the Flexicoking gasifier with the air feed, remains with the C_1-C_3 fraction after the cryogenic separation step, and subsequently enters the steam reformer and the methanator. In the ammonia synthesis section, the compressed hydrogen-rich gas from methanation is further cooled, and additional compression condensate is removed by physical separation employing a knockout drum (condensate removed as Stream 452) and with alumina drying agents. Ammonia synthesis typically proceeds at about 620°K (550°F) and 13 MPa (1900 psia) over an iron oxide catalyst (33). Ammonia is removed from the product gas stream by water scrubbing and sent to ammonia recovery for dehydration. The purified, high pressure hydrogen following nitrogen removal is sent to the liquefaction and solvent hydrogenation units as required.

In the Exxon design for the EDS MFS case, hydrogen production consists of a hydrogen recovery section and a hydrogen generation section. The cryogenic hydrogen recovery section for the MFS case is similar to that for the base case, except that less off-gas from the Flexicoking reactor is processed (since only half of the vacuum-bottoms slurry is being sent to Flexicoking) and the C_1/C_2 is recovered as high-Btu gas for sale and not as feed to the steam reformer.

Hydrogen generation for the MFS case involves syngas generation, shift conversion, acid gas removal and compression. Approximately half of the vacuum-bottoms slurry produced in the liquefaction/distillation units will be processed into hydrogen for the process while the balance will be processed through Flexicoking to produce additional liquid and gas products.

Production of raw synthesis gas for MFS hydrogen generation will be based upon coal gasification technology such as the Texaco partial oxidation process. The Texaco gasification process involves a pressurized, downflow, slagging gasifier which gasifies the vacuum-bottoms slurry with oxygen and steam. Pilot units gasifying coal operate at pressures of 2.1 to 8.2 MPa (300 to 1200 psia);

available test data and design information indicate that pressures in this range are appropriate for gasifying liquefaction residues similar to those generated by the EDS process (12,27). Gasification temperatures are generally above the ash fusion temperature (1500°K; 2300°F) to obtain high gasification rates and minimize the quantities of undesirable by-products such as tars, oils and phenols in the raw gas.

The gasifier is a refractory-lined carbon steel vessel which can roughly be divided into two zones, a gasification zone and a quenching zone. During gasification, the feed is partially reacted with oxygen, in the presence of steam, to produce a raw gas consisting primarily of CO, H_2, CO_2, and steam contaminated by H_2S and other reduced sulfur species (resulting from the largely inorganic sulfur in the vacuum bottoms), ammonia and other trace species. Quenching takes place in the lower portion of the reactor where the raw gas is partially cooled and the slagged ash is solidified through contact with water in a quench bath. The quenched gas is scrubbed with water to remove additional char and impurities such as ammonia and formate prior to subsequent processing (e.g., shift conversion). The quench water, after removal of the coarse slag and part of the slag fines by clarification, is combined with the used scrubber water. Most of this sour water is recycled as quench or scrubber water; however, a portion of the sour water is bled off, depressurized (Stream 441), and sent to wastewater treatment, in order to control the buildup of soluble ash constituents as well as organic and inorganic reaction products in the recirculating quench and scrubber water. When the blowdown stream is depressurized, a flash gas is derived (Stream 440) which is sent to the sulfur recovery area for processing.

Solids generated during gasification are slag and char. Quenched slag is removed from the gasifier through an ash lock system, and sized into coarse and fine fractions using a moving screen. Coarse slag is readily dewatered, while slag fines contained in a slurry (after separation from the quench water by clarification) require additional processing by thickening and filtration. Slag fines are filtered in the slag handling area, combined with the coarse slag and trucked to disposal. Filtrate from dewatering of the fine slag slurry (Stream 443) is pumped to wastewater treatment for suspended solids removal. Although not considered in the current EDS conceptual design, the small quantity of char containing approximately 6-12% unreacted carbon, may be recovered from the recirculating quench and scrubber water by settling. Depending upon the carbon content of the char, this material may be recycled to the gasifier, added to the coal feed in the front end of the liquefaction plant, or fed to the coal-fired boiler.

Carbon monoxide in the synthesis gas leaving the gasifier reacts with steam to produce hydrogen and carbon dioxide by combined high and low temperature shift conversion. High temperature shift (in range of 590-750°K; 600-890°F) proceeds in two stages over a cobalt-molybdenum catalyst. Gas leaving the high temperature shift reactor is cooled in a waste heat boiler and further shifted at low temperature (530-560°K; 490-550°F) over a cobalt-molybdenum catalyst. Both the high and low temperature shift reactions are conducted in the presence of hydrogen sulfide (hence, the term "sour shift") and sulfur-tolerant catalysts are used. The gases leaving the shift reactors will consist largely of H_2, CO_2, steam, CO and contaminants such as H_2S and ammonia. In the Exxon design, removal of acid gases from the shift gas employs the Catacarb process, as in the base case, although an alternative AGR process could be considered. A Catacarb or alternative AGR unit will remove most of the carbon dioxide, hydrogen sulfide, ammonia and carbonyl sulfide along with a small amount of hydrogen, carbon monoxide and methane. Acid gas from regeneration of the Catacarb (or alternative AGR) solution is sent to sulfur recovery. Purified hydrogen leaving the Catacarb absorber is cooled for condensation of water vapor (part of Stream 430), combined with the hydrogen from cryogenic recovery, compressed and sent to the liquefaction and solvent hydrogenation units as required.

Methanation, which is included in the EDS base case design to remove residual carbon oxides and prevent poisoning of the ammonia synthesis catalyst, has been deleted along with the ammonia synthesis section in the EDS MFS case design. In the base case design, ammonia synthesis is utilized to purge nitrogen from the hydrogen recovery/generation system. In the MFS case design, most of the nitrogen and carbon monoxide present in the feed stream to the hydrogen recovery section are removed along with the C_1/C_2 product gas stream in the cryogenic hydrogen separation step. Since most of the hydrogen for MFS process use is generated by partial oxidation of the vacuums-bottom slurry with steam and oxygen (no nitrogen introduced), the overall purity of the combined hydrogen from hydrogen recovery and partial oxidation is sufficiently high to be accepted for recycle to the liquefaction section. Thus, additional steps to remove any residual nitrogen in the MFS case are not necessary.

3.3.4.2.1 Gaseous Waste Streams

There are five gaseous waste streams from the hydrogen plant for the base case design, and six gaseous waste streams from the hydrogen plant for the MFS case design. Only one of these gaseous waste streams is common to both designs. These waste streams are:

- Stream 438 - deaerator vent gas from hydrogen generation
- Stream 426 - vent gas from CO_2 removal (base case only)
- Stream 434 - flue gas from reformer furnaces (base case only)
- Stream 446 - regeneration/decommissioning off-gas from the reformer catalyst (base case only)
- Stream 448 - decommissioning off-gas from the methanation catalyst (base case only)
- Stream 428 - acid gas from acid gas removal by Catacarb (MFS case only)
- Stream 440 - flash gas from the partial oxidation unit (MFS case only)
- Stream 449 - regeneration/decommissioning off-gas from the high temperature shift catalyst (MFS case only)

- Stream 450 - regeneration/decommissioning off-gas from the low temperature shift catalyst (MFS case only)
- Stream 802 - transient waste gas from partial oxidation unit (MFS case only).

Different gaseous waste streams are generated from the two EDS designs because of the major differences in the design of the hydrogen production plant. A less obvious case, however, is the absence of the regeneration/decommissioning off-gas from shift catalyst for the EDS base case design. This is because two different types of shift catalysts are used for the two EDS designs.

The high temperature and low temperature shift catalysts in the EDS MFS case design are sulfided cobalt-molybdenum catalysts because the shift reactions are carried out in the presence of hydrogen sulfide. These catalysts need to be regenerated for removal of sulfides, so that plant personnel can have safe access to shift converters for repairs and maintenance. Shift catalysts in the base case design (iron oxide catalyst) is not sulfided and generally not regenerated for process or safety reasons. Thus, there is no regeneration/decommissioning off-gas from shift catalyst for the base case design.

Deaerator vent gas from hydrogen generation (Stream 438) (Hydrogen production - Area 4)

Steam is needed for the steam reformer in the base case design and for the partial oxidation unit in the MFS case design. The feedwater to the boilers generating this steam includes in both cases condensed water from elsewhere in the plant, and therefore needs to be deaerated. The deaerator in the hydrogen production section serves two purposes: to increase boiler feedwater temperature by direct contact with low pressure exhaust steam, and to remove most undesirable dissolved gases from boiler feedwater due to its prior use. Stream 438 is the small amount of steam (about 1% of the boiler feedwater supplied) and non-condensible gases that escape through the deaerator vent. For the base case design, the boiler feedwater to the deaerator (to generate steam for the steam reformer) includes approximately 39% condensate from the high temperature shift gas and 61% demineralized water. For the MFS case design, the boiler feedwater to the deaerator (to generate steam for the partial oxidation unit) includes approximately equal portions of condensate from the low temperature shift gas and demineralized make-up water. Thus, the deaerator vent gas is expected to contain traces of shift gas components that are dissolved in the condensates. The deaerator vent gas should consist primarily of water vapor with a small quantity of carbon dioxide and traces of light hydrocarbons in the base case, and additional H_2S and NH_3 in the MFS case, although detailed characterization data are not available. For both the base case and MFS case designs, the deaerator vent gas is also estimated to contain 0.6 ppmv CO, with total CO emissions of less than 0.5 kg/hr (20). Assuming that the flow rate of this vent gas is approximately equal to 1% of the boiler feedwater supplied to the deaerator, the deaerator vent gas is generated at the rate of 6.5 kmol/hr for the base case design and 6.2 kmol/hr for the MFS case design.

Vent gas from CO_2 removal (Stream 426) (Hydrogen production - Area 4)

The Catacarb process is used for the bulk removal of carbon dioxide from the gas leaving the shift reactor in the base case design, although alternative AGR systems could be considered. In the Catacarb absorber, most of the carbon dioxide and water vapor in the shift gas is removed along with a small amount of hydrogen, carbon monoxide, and methane. Vent gas (Stream 426) is generated from the regeneration of the Catacarb solution (found only in the base case design). Stream 426 will consist primarily of carbon dioxide and water vapor, along with 1.9% hydrogen, 370 ppmv methane and 160 ppmv carbon monoxide. Any ammonia which may be generated by reaction of hydrogen and nitrogen during hydrogen generation would also be present in the vent gas. Characterization data for this waste stream are presented in Table 3-42.

TABLE 3-42. ESTIMATED COMPOSITION AND FLOW RATES OF VENT GAS FROM CO_2 REMOVAL (STREAM 426)*

Component	Base Case
H_2, kmol/hr	118.1
C_1, kmol/hr	2.2
N_2, kmol/hr	0
CO, kmol/hr	0.96
CO_2, kmol/hr	4,782.7
H_2S, kmol/hr	0
COS, kmol/hr	0
H_2O, kmol/hr	1,196.0
Total, kmol/hr	6,100.0
Temperature, °K	339
Pressure, MPa	0.1

Data source: Data for CO_2 and H_2 are obtained from Attachment 3 of Reference 20, composition of other components is estimated based upon available data for hot carbonate process in Reference 28.

* Stream 426 is generated only from the base case design.

Liquefaction Residue Processing/
Hydrogen Production - Area 4
Streams 434, 446

Flue gas from reformer furnaces (Stream 434) (Hydrogen production - Area 4)

This flue gas stream is generated from supplying heat to the steam reformers by the combustion of the low-Btu fuel gas provided from the Flexicoking gas, and applies only to the base case design. Total heat provided by the fuel gas is 3.165 TJ/hr (3000 MM Btu/hr). The low-Btu fuel gas contains 80 ppmv H_2S and 3 ppmv COS (13). Assuming combustion with 20% excess air, the flow rate of the flue gas generated was calculated to be 58,023 kmol/hr (808,000 SCFM). In Table 3-43, the emission factors used and the estimated emission rates of pollutants for this flue gas stream are presented.

Regeneration/decommissioning off-gas from the reformer catalyst (Stream 446) (Hydrogen production - Area 4)

In steam reforming, carbonaceous deposits are gradually formed on the surface of the nickel-urania catalyst. The continuing accumulation of the carbonaceous deposits reduces the activity of the catalyst to the point that it must be regenerated. It is assumed that once a year on the average, the reformer catalyst will be regenerated by a mixture of steam and air. Thus, the regeneration/decommissioning off-gas from the nickel-urania reformer catalyst will be intermittent in nature, and applies only to the base case design. No data are available regarding the composition of this off-gas; however, it is anticipated to consist primarily of steam and nitrogen with small amounts of CO and particulate matter because of the steam-carbon reaction and catalyst carryover. Small quantities of $Ni(CO)_4$ may also be present in the off-gas. Estimated generation rate of this off-gas, based on typical catalyst regeneration off-gas values (29), is 6,800 kmol/hr. The average emission frequency for this waste stream is once a year, with an average emission duration of 24 hours.

TABLE 3-43. ESTIMATED UNCONTROLLED EMISSIONS FOR FLUE GAS FROM REFORMER FURNACES (STREAM 434)*

Pollutant	Emission Factor † ng/J (lb/MM Btu)	Emission Rate kg/hr
Particulate Matter	2.1 - 6.4 (0.005-0.015)	6.8 - 20.4
NO_x (as NO_2)	52 - 99 (0.12-0.23)	163 - 313
CO	7.3 - 10.7 (0.017-0.025)	23 - 34
Hydrocarbons	1.3 (0.003)	4.1
SO_2	48 (0.11)	150

* Stream 434 is only generated from the base case design.

† Emission factors for particulate matter, NO_x, and hydrocarbons are obtained from AP-42 for natural gas combustion in industrial process heaters (21). By comparison, Exxon estimated emission factors of 3.1 ng/J for particulate matter, and 60 ng/J for NO_x (20). The emission factor for CO is based on 7.3 ng/J from AP-42 and 10.7 ng/J from Exxon. The emission factor for SO_2 is based on 83 ppmv total reduced sulfur in the low-Btu fuel gas (13).

Decommissioning off-gas from the methanation catalyst (Stream 448) (Hydrogen production - Area 4)

The nickel catalyst used in methanation is pyrophoric and may lead to spontaneous ignition in the reduced state. Prior to disposal, the nickel catalyst would be oxidized by air, with steam added for temperature control. It is assumed that methanation catalysts will be decommissioned once every four years. Thus, the decommissioning off-gas from the methanation catalyst will be intermittent in nature. This stream is found only in the base case design. Based upon composition of the feed gas to the methanation unit and typical catalyst decommissioning procedures, this off-gas is expected to consist primarily of steam and nitrogen with small amounts of particulate matter, and free of sulfur and carbon oxides. This is because sulfur species have been removed prior to methanation and carbon deposition on catalysts is not known to be a problem in methanation. Although there are no supporting data, small quantities of $Ni(CO)_4$ may also be present in the off-gas. Using typical catalyst regeneration off-gas flow rate values (29), Stream 448 was estimated to be generated at the rate of approximately 5,000 kmol/hr. The average emission frequency of this waste stream is once every four years, with an average emission duration of 24 hours.

Acid gas from acid gas removal (Stream 428) (Hydrogen production - Area 4)

The Catacarb process is used for the removal of acid gases from the shift gas in the MFS design, although alternative acid gas removal techniques might also be considered for this application. Because the H_2-rich syngas produced by partial oxidation of the sulfur- and nitrogen-containing vacuum bottoms slurry is sent to the shift reactors without extensive gas cleaning, the shift gas will contain H_2S and NH_3 as well as CO_2. This gaseous waste stream results from regeneration of the Catacarb solution, and applies only to the MFS case design. The estimated composition of the shift gas that is fed to the Catacarb unit is presented in Table 3-44. These data may be used to evaluate alternative

TABLE 3-44. ESTIMATED COMPOSITION AND FLOW RATES OF SHIFT GAS TO THE ACID GAS REMOVAL UNIT IN MFS CASE DESIGN

Component	Shift Gas to Acid Gas Removal*
H_2, kmol/hr	18,506.7
C_1, kmol/hr	90.0
N_2, kmol/hr	122.3
CO, kmol/hr	263.2
CO_2, kmol/hr	12,392.9
H_2S, kmol/hr	137.4
NH_3, kmol/hr	8.9
COS, kmol/hr	0.6
H_2O, kmol/hr	1,060.6
HCN, kmol/hr	0.02
Total, kmol/hr	3,258.6
Temperature, °K	394
Pressure, MPa	6.2

Data source: Estimated by using test and design data from References 12, 20, 26 and 27.

* Composition of the shift gas after cooling to 394 °K and condensation of water vapor present in the hot shift gas.

acid gas removal (AGR) systems. Assuming the Catacarb process is the AGR technique employed, the composition of the acid gas stream from regeneration of the Catacarb solution has been estimated based upon Exxon design estimates and test data from tests by others on the Catacarb system (20,26,27,28). The estimated characteristics of this acid gas stream are presented in Table 3-45. These estimated characteristics show that Stream 428 is expected to contain 1% H_2S, 670 ppmv NH_3, 60 ppmv CO, and 40 ppmv COS.

Flash gas from the partial oxidation unit (Stream 440) (Hydrogen production - Area 4)

Flash gas from the partial oxidation unit is generated on a continuous basis from the depressurization of the bleed stream in the recirculating quench and scrubber water loop. The dissolved gases (H_2S, NH_3, H_2, CO, CO_2, COS and CH_4) build up at the elevated gasifier pressure in the recirculating water loop, and are partially released when the bleed stream is depressurized. Assuming that the Texaco gasifier is employed for the partial oxidation of the vacuum-bottoms slurry, the flash gas will consist primarily of H_2, CO, CO_2 with 4.4% H_2S, 0.12% COS, and 150 ppmv NH_3 (27). Composition estimates for this flash gas stream, as presented in Table 3-46, are based upon Texaco pilot plant tests gasifying SRC-II vacuum bottoms from Kentucky 9/14 coal. No characterization data for this stream using EDS vacuum-bottoms slurry and Illinois No. 6 coal are publicly available as of this writing.

Regeneration/decommissioning off-gas from the high temperature shift catalyst (Stream 449) (Hydrogen production - Area 4)

As discussed previously in the introduction to this section, shift catalysts in the MFS case design are of the sulfided form. These catalysts require periodic regeneration for removal of deposited carbon and dusts. Sulfide removal is also necessary to provide safe access to the shift converters for repairs and maintenance. The shift units in an EDS plant would be modular with several parallel trains. During regeneration, one or more trains would be

TABLE 3-45. ESTIMATED COMPOSITION AND FLOW RATES OF ACID GAS FROM ACID GAS REMOVAL UNIT (STREAM 428)*

Component	Stream 428 MFS Case
H_2, kmol/hr	112.9
C_1, kmol/hr	0.3
CO, kmol/hr	0.8
CO_2, kmol/hr	12,130.1
H_2S, kmol/hr	137.1
NH_3, kmol/hr	8.9
COS, kmol/hr	0.5
H_2O, kmol/hr	913.0
Total, kmol/hr	13,303.6
Temperature, °K	316
Pressure, MPa	0.1

Data source: CO_2, H_2S and H_2O composition from Attachment 4 of Reference 20, composition of other components estimated using shift gas composition (Table 3-44) and Catacarb process model (28).

* Stream 428 is only generated from the MFS case design.

TABLE 3-46. ESTIMATED COMPOSITION AND FLOW RATES OF FLASH GAS FROM PARTIAL OXIDATION UNIT (STREAM 440)*

Component	Stream 440 MFS Case
H_2, kmol/hr	61.3
C_1, kmol/hr	0.08
N_2, kmol/hr	1.4
CO, kmol/hr	77.7
CO_2, kmol/hr	82.7
H_2S, kmol/hr	11.2
NH_3, kmol/hr	0.04
COS, kmol/hr	0.3
H_2O, kmol/hr	19.0
Total, kmol/hr	253.7
Temperature, °K	322
Pressure, MPa	0.2

Data source: Pilot plant tests performed with SRC-II residue using Kentucky 9/14 coal (27).

* Stream 440 is only generated from the MFS case design.

taken off line for regeneration while others remain in service. It is assumed that regeneration of each train of high temperature and low temperature shift catalysts will be conducted once a year. Regeneration is by reaction with steam and air, driving off SO_2. Thus, emissions from regeneration would be intermittent in nature, depending on the number of trains and the exact regeneration schedule. The duration of regeneration may range from 12 hours to 72 hours, depending on which shift stage is being regenerated. Off-gas from regeneration/decommissioning of high and low temperature shift catalyst (Streams 449 and 450) would consist primarily of steam with 6% N_2, 1% SO_2 and 0.5% CO_2 (29). Characterization data for Stream 449 and the regeneration/decommissioning off-gas from low temperature shift catalyst (Stream 450) are presented in Table 3-47. Data on the characteristics of the individual waste streams are not available. The average emission frequency for Streams 449 and 450 is once a year, with an average emission duration of 24 hours.

Regeneration/decommissioning off-gas from the low temperature shift catalyst (Stream 450) (Hydrogen production - Area 4)

The characteristics of this waste stream have been discussed in conjunction with Stream 449.

Transient waste gas from partial oxidation units (Stream 802) (Hydrogen production - Area 4)

Transient waste gas is generated from the partial oxidation units in the MFS case design. It is assumed that transient conditions generating waste gases would be expected only once a year, and only to one of the units and up to a day in duration (13). The transient waste gas from the partial oxidation unit is anticipated to be similar in composition to the quenched syngas from the unit. Estimated characteristics for this waste stream, based upon EDS design data and data for the Texaco gasifier when processing SRC-II vacuum flash drum bottoms from Kentucky No. 9/14 coal (20,27), are presented in Table 3-48. The estimated flow rate for this waste stream is 12,120 kmol/hr.

TABLE 3-47. ESTIMATED COMPOSITION AND FLOW RATES OF REGENERATION/DECOMMISSIONING OFF-GAS FROM SHIFT CATALYST (STREAMS 449 AND 450)*

Component	Streams 449/450 MFS Case
N_2, kmol/hr	1,357.6
CO_2, kmol/hr	111.3
SO_2, kmol/hr	244.8
H_2O, kmol/hr	20,540.9
Total, kmol/hr	22,254.6
Particulate matter, kg/hr	608
Temperature, °K	No data
Pressure, MPa	0.1

Data source: Off-gas composition is based upon estimates provided in permit applications for the ANR Synthetic Natural Gas Plant (29)

* Streams 449/450 are generated only from the MFS case design.

Intermittent stream: average occurrence of once a year, 24 hours per occurrence.

TABLE 3-48. ESTIMATED CHARACTERISTICS OF TRANSIENT WASTE GAS FROM PARTIAL OXIDATION UNITS (STREAM 802)*

Component	Volume Percent in Waste Gas	Flow Rate, kmol/hr
H_2	20.2	2,454.5
C_1	0.2	24.3
CO	17.7	2,150.7
CO_2	7.9	959.9
H_2S	0.3	36.5
NH_3	170 ppmv	2.1
COS	250 ppmv	3.1
H_2O	53.4	6,488.6
HCN	8 ppmv	0.1
HCO_2H	70 ppmv	0.9
Total flow rate		12,120.7
Temperature, °K		No data
Pressure, MPa		No data

Data source: Estimated from EDS design data (20) and data for the Texaco gasifier when processing SRC-II vacuum flash drum bottoms from Kentucky No. 9/14 coal (27).

* Stream 802 is generated only from the MFS case design.

Intermittent stream: average occurrence of once a year, and up to 24 hours per occurrence.

3.3.4.2.2 Liquid Waste Streams

There are five aqueous waste streams from the hydrogen plant for the base case design, and four aqueous waste streams from the hydrogen plant for the MFS case design. Only two of the aqueous waste streams are common to both designs. These waste streams are:

- Stream 403 - knockout drum wastewater from cryogenic hydrogen recovery
- Stream 430 - blowdown and knockout drum wastewater from hydrogen generation
- Stream 431 - Catacarb overhead receiver wastewater (base case only)
- Stream 451 - aqueous ammonia from ammonia synthesis (base case only)
- Stream 452 - knockout drum wastewater from ammonia synthesis (base case only)
- Stream 441 - sour water from the partial oxidation unit (MFS case only)
- Stream 443 - slag filtrate from the partial oxidation unit (MFS case only).

Knockout drum wastewater from cryogenic hydrogen recovery (Stream 403) (Hydrogen production - Area 4)

For both the base case and the MFS case designs, the cryogenic hydrogen recovery feed is comprised of hydrogen-rich purge gases from the high-pressure liquefaction (Area 2) and solvent hydrogenation (Area 3) separators, and off-gases from the low-pressure atmospheric fractionator distillate drum (Stream 151, Area 3), solvent stripper (Stream 251, Area 3), and Flexicoking unit reactor system (Stream 310, Area 4). The off-gases are treated by DEA scrubbing, compressed to 5.2 MPa (740 psig), and cooled to 297°K (75°F). These off-gases will contain moisture because of moisture and oxygen present in coal feed, excess steam added to the Flexicoking reactor, and possibly water evaporated from wash water and scrubbing solutions in the processing sequence.

The cooling sequence condenses out most of the moisture in knockout drums as Stream 403. Detailed characterization data for this wastewater stream are not available, although traces of hydrogen sulfide, ammonia, carbon dioxide, phenols, and organic acids may be present, as shown in Table 3-49. The estimated generation of Stream 403 is 1,334 kg/hr (1.3 m^3/hr) for both the base case and the MFS case design.

Blowdown and knockout drum wastewater from hydrogen generation (Stream 430) (Hydrogen production - Area 4)

Steam is used as feed to the steam reformer in the base case design and to the partial oxidation unit in the MFS case design. The blowdown from the steam drum of the boilers providing this steam constitutes approximately 20% of Stream 430 in the base case and 35% of Stream 430 in the MFS case. In the base case, hydrogen-rich gas leaving the methanator is compressed and cooled to condense out most of the water vapor. The water vapor results from steam fed to the steam reformer that remains unreacted after the shift conversion. The condensate from the methanator effluent constitutes the remaining 80% of Stream 430 in the base case. In the MFS case, hydrogen-rich gas leaving the Catacarb absorber is further cooled to 317°K (110°F) to condense out water vapor, and this condensate constitutes the remaining 65% of MFS Stream 430. The water vapor in this case results from steam fed to the partial oxidation unit that remains unreacted after the shift conversion. For both designs, Stream 430 should be relatively free of contaminants because boiler feedwater consists of deaerated demineralized water and shift condensates but detailed characterization data are not available. Estimated characteristics of Stream 430 are presented in Table 3-50. Although not shown in the table, the stream should also contain total dissolved solids (TDS) and boiler feedwater additives for scale and corrosion control such as hydrazine and phosphates. The generation rate of this wastewater stream is 55,430 kg/hr (57.7 m^3/hr) for the base case design and 3,110 kg/hr (3.2 m^3/hr) for the MFS case design. The differences in the generation rates are due to the differences in the steam

TABLE 3-49. ESTIMATED QUALITY OF KNOCKOUT DRUM WASTEWATER FROM CRYOGENIC HYDROGEN RECOVERY (STREAM 403)

Contaminant	Concentration ppmw	Flow Rate, kg/hr
H_2S	Trace	0
NH_3	Trace	0
HCl	Trace	0
CO_2	Trace	0
Phenols	Trace	0
Organic Acids	Trace	0
Total flow rate, kg/hr		1,334
Total flow rate, m^3/hr		1.3
Temperature, °K		316
Pressure, MPa		10.3

Data source: Table 1-X-1 of Reference 13.

TABLE 3-50. ESTIMATED QUALITY OF BLOWDOWN AND KNOCKOUT DRUM WASTEWATER FROM HYDROGEN GENERATION (STREAM 430)

Contaminant*	Base Case		MFS Case	
	Concentration, ppmw	Flow Rate kg/hr	Concentration ppmw	Flow Rate kg/hr
H_2S	0	0	Trace	0
NH_3	0	0	Trace	0
HCl	0	0	0	0
CO_2	Trace	0	Trace	0
Phenols	0	0	0	0
Organic Acids	0	0	0	0
Total flow rate, kg/hr		55,430		3,110
Total flow rate, m^3/hr		57.7		3.2
Temperature, °K		366		366
Pressure, MPa		1.7		1.7

Data source: Composition and flow rate data for the base case design were obtained from Table 1-X-1 of Reference 13. Composition data for the MFS case design are slightly different from base case design because sour shift is employed in the MFS case. Flow rate for the MFS case was calculated using data presented in Tables 3-44, 3-45 of this section and Figure 2-VII-2 of Reference 13.

*Other contaminants present should include total dissolved solids (TDS) and boiler feedwater additives for scale and corrosion control.

requirements and the amount of water vapor present in the effluent from the shift conversion units.

Catacarb overhead receiver wastewater (Stream 431) (Hydrogen production - Area 4)

This wastewater stream is generated from condensation of water vapor present in the overhead gas grom the Catacarb regenerator, and applies only to the base case design. For the MFS case design, this wastewater stream is not generated because there is less water vapor present in the feed gas to the MFS Catacarb absorber, and any water subsequently condensed in the regenerator overhead is combined with the regenerated Catacarb solution and returned to the absorber. Estimated characteristics of this stream are presented in Table 3-51. Stream 431 should be free of hydrogen sulfide, phenols and organic acids, since the base case Catacarb is simply sorbing CO_2 from a clean hydrogen stream. However, Stream 431 will contain 600 ppmw CO_2 and may contain traces of ammonia. The generation rate of this waste stream is 22,690 kg/hr (23.2 m^3/hr).

Aqueous ammonia from ammonia synthesis (Stream 451) (Hydrogen production - Area 4)

The ammonia synthesis section removes nitrogen from the hydrogen by catalytic conversion over iron oxide to ammonia. The ammonia synthesis reactor effluent is cooled and sent to a water scrubber where aqueous ammonia is recovered as Stream 451 in the base case design. Estimated characteristics of this stream are presented in Table 3-52. The aqueous ammonia contains 11.5% NH_3 and possibly traces of H_2S and organics. The generation rate of this stream is 22,560 kg/hr (24.1 m^3/hr). The ammonia present in this stream is a potentially salable by-product.

Knockout drum wastewater from ammonia synthesis (Stream 452) (Hydrogen production - Area 4)

This wastewater stream is generated from the cooling and condensation of residual water vapor present in the hydrogen-rich gas which is fed to the ammonia synthesis reactor for nitrogen removal, and applies only to the base

TABLE 3-51. ESTIMATED QUALITY OF CATACARB OVERHEAD RECEIVER WASTEWATER (STREAM 431)*

Contaminant	Concentration, ppmw	Flow Rate, kg/hr
H_2S	0	0
NH_3	Trace	0
HCl	0	0
CO_2	600	13.6
Phenols	0	0
Organic Acids	0	0
Total flow rate, kg/hr		22,690
Total flow rate, m^3/hr		23.2
Temperature, °K		339
Pressure, MPa		1.4

Data source: Table 1-X-1 of Reference 13.

* Stream 431 is generated only from the base case design.

TABLE 3-52. ESTIMATED QUALITY OF AQUEOUS AMMONIA FROM AMMONIA SYNTHESIS (STREAM 451)*

Contaminant	Concentration ppmw	Flow Rate, kg/hr
H_2S	Trace	0
NH_3	115,260	2,600
HCl	0	0
CO_2	Trace	0
Phenols	0	0
Organic Acids	0	0
Total flow rate, kg/hr		22,560
Total flow rate, m^3/hr		24.1
Temperature, °K		322
Pressure, MPa		12.9

Data source: Table 1-X-1 of Reference 13.

* Stream 451 is generated only from the base case design.

case design. Any pollutants present in Stream 452 are expected to be at trace levels since the hydrogen-rich gas from which the water condenses will be fairly clean before it reaches the ammonia synthesis step. Estimated characteristics of this stream are presented in Table 3-53. The generation rate of this wastewater stream is 450 kg/hr (0.45 m^3/hr).

Sour water from the partial oxidation unit (Stream 441) (Hydrogen production - Area 4)

The gas stream leaving the gasification zone of the Texaco gasifier is quenched with water to remove most of the slagged ash, and the quenched gas is scrubbed with water to remove additional char and impurities such as ammonia and formate. The quench water, after removal of the coarse slag and part of the slag fines by clarification, is combined with the used scrubber water. Most of this sour water is recycled as quench or scrubber water; however, a portion of the sour water is bled off in order to control the buildup of soluble ash consitutents as well as organic and inorganic reaction products in the recirculating quench and scrubber water. This bleed stream, after depressurization, is Stream 441. Thus, Stream 441 originates from the condensation of the water vapor present in the effluent from the partial oxidation unit as a result of steam injected into this gasifier. Estimated characteristics of Stream 441, based upon test data and design data for gasification of various coals in a Texaco gasifier (12,30,31,32), are presented in Table 3-54. The stream is expected to contain 65-336 ppmw H_2S, 1220-2700 ppmw NH_3, 492-1200 ppmw formate, and very low levels of phenols. Also, based upon very limited test data on coal gasification in a Texaco coal gasifier, polycyclic hydrocarbons (e.g., anthracene, pyrene) are present at parts per billion levels (12). In Table 3-55, the concentrations of trace elements in Stream 441 are presented, also based on coal gasification tests. It is emphasized that the numbers in Tables 3-54 and 3-55 are based upon tests gasifying various coals and could be different from compositions when gasifying EDS vacuum bottoms resulting from Illinois No. 6 coal. When ranges of concentration values are presented, the

TABLE 3-53. ESTIMATED QUALITY OF KNOCKOUT DRUM WASTEWATER IN AMMONIA SYNTHESIS (STREAM 452)*

Contaminant	Concentration, ppmw	Flow Rate, kg/hr
H_2S	Trace	0
NH_3	Trace	0
HCl	0	0
CO_2	Trace	0
Phenols	0	0
Organic Acids	0	0
Total flow rate, kg/hr		454
Total flow rate, m^3/hr		0.45
Temperature, °K		311
Pressure, MPa		12.4

Data source: Table 1-X-1 of Reference 13.

* Stream 452 is generated only from the base case design.

TABLE 3-54. ESTIMATED QUALITY OF SOUR WATER FROM THE PARTIAL OXIDATION UNIT (STREAM 441)*

Contaminant	Concentration, ppmw	Flow Rate, kg/hr
H_2S	65-336	3-15
NH_3	1220-2700	56-123
HCl	No data	No data
CO_2	No data	No data
Phenols	0.023	0.001
Organic Acids	No data	No data
Formate	492-1200	22-55
Sulfate	21-166	1.0-7.6
Sulfite	180	8.2
$S_2O_3^=$	69	3.1
Cl^-	7.5-1320	0.3-60
CN^-	8-45	0.4-2.0
SCN^-	8-70	0.4-3.2
COD	405-760	18-35
TDS	420-2000	19-91
TSS	330	15
TOC	215-760	9.8-35
pH	7.7-8.8	Not applicable
Total flow rate, kg/hr		45,550
Total flow rate, m^3/hr		45.5
Temperature, °K		355
Pressure, MPa		0.14

Data source: H_2S, CN^-, Cl^-, TDS, and TOC data were obtained from References 12, 30 and 31; NH_3 and formate data were obtained from References 12, 30, 31 and 32; phenol and sulfite data were obtained Reference 31; sulfate and SCN- data were obtained from References 12, 31, and 32; $S_2O_3^=$ data were obtained from Reference 32; COD and pH data were obtained from References 12 and 31; TSS data were obtained from Reference 31. Flow rate was extrapolated from the SRC-II design for the Texaco gasifier (26). Temperature and pressure were obtained from Table 2-VII-2 of Reference 13. Data from References 12, 30, 31 and 32 were acquired in the gasification of both eastern and western coals in pilot Texaco units.

* Stream 441 is generated only from the MFS case design.

TABLE 3-55. CONCENTRATIONS OF TRACE ELEMENTS IN SOUR WATER FROM THE PARTIAL OXIDATION UNIT (STREAM 441)*

Element		Concentration, ppmw
Aluminum	Al	20
Antimony	Sb	<0.001
Arsenic	As	0.022
Barium	Ba	0.54
Beryllium	Be	0.018
Bromine	Br	<1
Cadmium	Cd	0.003
Calcium	Ca	140
Chromium	Cr	<0.003
Cobalt	Co	0.021
Copper	Cu	0.01
Fluorine	F	39-500
Iron	Fe	3.7
Lead	Pb	0.054
Magnesium	Mg	100
Manganese	Mn	0.10
Mercury	Hg	<0.0001
Molybdenum	Mo	<0.02
Nickel	Ni	0.03
Selenium	Se	0.010
Silicon	Si	5.0
Silver	Ag	0.002
Sodium	Na	80
Thallium	Tl	0.01
Vanadium	V	<0.04
Zinc	Zn	0.03

Data source: Data for Al, Ca, Fe, Mg, Na and Si were obtained from Reference 32; data for Br were obtained from References 12 and 31; data for F were obtained from References 12, 31 and 32; all other data were obtained from Reference 12. Data from these references were acquired in the gasification of both eastern and western coals in pilot Texaco units.

* Stream 441 is generated only from the MFS case design.

higher value will be used in the PCTM evaluation of wastewater treatment technology. The generation rate of Stream 441, estimated by extrapolating the SRC-II commercial plant design data for processing SRC-II vacuum bottoms in the Texaco gasifier (26), is 45,550 kg/hr (45.5 m^3/hr). The composition/flow estimates from Stream 441 were prepared using coal gasification data and SRC-II design estimates, because no data/estimates are publicly available addressing gasification of vacuum bottoms from the EDS design facility.

Slag filtrate from the partial oxidation unit (Stream 443) (Hydrogen production - Area 4)

Most of the slag fines in the quench water from the Texaco gasifier are removed by clarification and a slag fines slurry is obtained. Stream 443 is the wastewater stream that results from filtration of the slag fines slurry. Estimated characteristics are based upon test data for the Texaco coal gasifier (12,31) and presented in Table 3-56. The data show that the concentration levels of contaminants in the slag filtrate are generally an order of magnitude lower than the concentration levels of the same contaminants in the sour water from the partial oxidation unit. The concentrations of trace elements in Stream 443 are presented in Table 3-57. As with the sour water from the partial oxidation unit, the numbers in Tables 3-56 and 3-57 are based upon results from gasifying a variety of coals, not EDS vacuum bottoms; the results from gasification of vacuum bottoms could be different. In Tables 3-56 and 3-57, the higher value will be used in the evaluation of wastewater treatment technology when ranges of concentration values are presented. The generation rate of Stream 443 was estimated by extrapolating SRC-II design estimates to be 9,320 kg/hr (9.3 m^3/hr). No data are available from Exxon concerning the characteristics of this wastewater stream.

TABLE 3-56. ESTIMATED QUALITY OF SLAG FILTRATE FROM THE PARTIAL OXIDATION UNIT (STREAM 443)*

Contaminant	Concentration, ppmw	Flow Rate, kg/hr
H_2S	<0.5-6	<0.005-0.06
NH_3	130-380	1.2 - 3.5
HCl	No data	No data
CO_2	No data	No data
Phenols	<0.01	<0.0001
Organic Acids	No data	No data
Formate	56-94	0.52-0.88
Sulfate	5-38	0.05-0.35
Sulfite	4.7	0.044
$S_2O_3^=$	5	0.05
Cl^-	42-146	0.39-1.4
CN^-	0.4-1	0.004-0.01
SCN^-	2-4	0.019-0.037
COD	48-221	0.45-2.1
TDS	28-355	0.26-3.3
TSS	No data	No data
TOC	28-38	0.26-0.35
pH	8-8.9	
Total flow rate, kg/hr		9,320
Total flow rate, m^3/hr		9.3
Temperature, °K		303
Pressure, MPa		0.10

Data source: Data for sulfite were obtained from Reference 31; data for S_2O_3 were obtained from Reference 12; all other concentration data were obtained from References 12 and 31. Flow rate was extrapolated from the SRC-II design for the Texaco gasifier (26). Temperature and pressure were obtained from Table 2-VII-2 of Reference 13. Data from References 12 and 31 were acquired in the gasification of both eastern and western coals in pilot Texaco units.

* Stream 443 is generated only from the MFS case design.

TABLE 3-57. CONCENTRATIONS OF TRACE ELEMENTS IN SLAG FILTRATE FROM THE PARTIAL OXIDATION UNIT (STREAM 443)*

Element		Concentration, ppmw
Aluminum	Al	No data
Antimony	Sb	<0.005-0.016
Arsenic	As	0.07-0.08
Barium	Ba	0.10-0.27
Beryllium	Be	<0.005
Bromine	Br	<1
Cadmium	Cd	<0.005
Calcium	Ca	No data
Chromium	Cr	<0.003
Cobalt	Co	<0.005-0.10
Copper	Cu	0.017-0.043
Fluorine	F	<0.1-31
Iron	Fe	No data
Lead	Pb	0.045-0.059
Magnesium	Mg	No data
Manganese	Mn	0.016-0.08
Mercury	Hg	<0.0001-0.11
Molybdenum	Mo	0.016-0.03
Nickel	Ni	<0.05-0.07
Selenium	Se	0.033-0.14
Silicon	Si	No data
Silver	Ag	0.002-0.01
Sodium	Na	No data
Thallium	Tl	<0.01
Vanadium	V	<0.05
Zinc	Zn	<0.05

Data source: References 12 and 31 for all elements. Data from References 12 and 31 were acquired in the gasification of eastern and western coals in pilot Texaco units.

* Stream 443 is generated only from the MFS case design.

3.3.4.2.3 Solid Waste Streams

There are eight solid waste streams from hydrogen production for the base case design, and five solid waste streams from hydrogen production for the MFS case design. Only two of the solid waste streams are common to both designs.

These solid waste streams are:

- Stream 404 - spent hydrotreater catalyst from cryogenic hydrogen recovery
- Stream 405 - spent drying agents from cryogenic hydrogen recovery
- Stream 433 - spent sulfur guard from hydrogen generation (base case only)
- Stream 435 - spent reformer catalyst from hydrogen generation (base case only)
- Stream 436 - spent shift catalyst from hydrogen generation (base case only)
- Stream 439 - spent methanation catalyst from hydrogen generation (base case only)
- Stream 453 - spent drying agents from ammonia synthesis (base case only)
- Stream 454 - spent ammonia synthesis catalyst (base case only)
- Stream 442 - slag from the partial oxidation unit (MFS case only)
- Stream 444 - spent high temperature shift catalyst from hydrogen generation (MFS case only)
- Stream 445 - spent low temperature shift catalyst from hydrogen generation (MFS case only).

Spent hydrotreater catalyst from cryogenic hydrogen recovery (Stream 404) (Hydrogen production - Area 4)

Hydrotreating over nickel-molybdenum catalyst is employed to remove olefins present in the off-gas from the Flexicoking reactor, prior to sending this gas stream to cryogenic separation for hydrogen recovery. The hydrotreater catalyst will require periodic replacement as a portion of it becomes deactivated as a result of buildup of carbonaceous deposits on the catalyst surface.

Based on Exxon's estimates (13), the consumption rate for the nickel-molybdenum catalyst is 65.8 Mg/yr for the base case design, and 31.8 Mg/yr for the MFS case design. Consumption rate for the inert balls is 2.7 Mg/yr and 5.4 Mg/yr for the base case and MFS case designs, respectively. Thus, approximately 68.5 Mg/yr of spent hydrotreater catalyst are generated for the base case design, and 37.2 Mg/yr for the MFS case design. The composition for the spent catalyst is similar to the fresh catalyst, and is expected to contain nickel, molybdenum, carbonaceous material, and sulfides. Leaching characteristics for this spent catalyst are not known.

Spent drying agents from cryogenic hydrogen recovery (Stream 405) (Hydrogen production - Area 4)

The gas stream entering the cold box in cryogenic hydrogen recovery is dried with zeolites to remove the remaining water, H_2S, and CO_2. The estimated generation rate for the spent zeolites used in cryogenic hydrogen recovery is 263 Mg/yr and 177 Mg/yr for the base case (6) and MFS case (8) designs, respectively. These spent molecular sieves are saturated with water, H_2S and CO_2, but detailed characterization data are not available.

Spent sulfur guard from hydrogen generation (Stream 433) (Hydrogen production - Area 4)

Zinc oxide sulfur guards are used to remove residual amounts of sulfur species present in the feed gas to the steam reformer, in order to protect reformer catalysts from sulfur poisoning. These sulfur guards must be replaced periodically as most of the zinc oxide has been reacted to form zinc sulfide. Stream 433 results from the generation of the spent sulfur guards, and applies only to the base case design. The estimated generation rate for the spent zinc oxide is 271 Mg/yr, including 38 Mg/yr of inert balls (13). The spent sulfur guard is expected to contain mostly zinc oxide, zinc sulfide, and carbonaceous material. Leaching characteristics for this solid waste are not known.

Liquefaction Residue Processing/
Hydrogen Production - Area 4
Streams 435, 436, 439

Spent reformer catalyst from hydrogen generation (Stream 435) (Hydrogen production - Area 4)

Steam reforming of light hydrocarbons over nickel-urania catalyst is employed for hydrogen production in the base case design. The reformer catalyst will require periodic replacement as some of it becomes deactivated as a result of fouling by carbon deposition or poisoning by sulfur compounds or other trace species. Based on Exxon's estimate of the catalyst consumption rate, the spent reformer catalyst is generated at the rate of 181 Mg/yr (13). Detailed characterization data for the spent reformer catalyst are not available. A representative composition for the fresh nickel-urania catalyst is 13.0 wt % Ni, 12.1 wt % U, 0.3 wt % K, and the balance supporting alumina (33). The composition for the spent catalyst is similar, but with the addition of carbonaceous material and traces of sulfides.

Spent shift catalyst from hydrogen generation (Stream 436) (Hydrogen production - Area 4)

For the base case design, the shift reaction (sweet shift) takes place after the removal of all sulfur species in the DEA acid gas removal unit and the zinc oxide sulfur guards located upstream of the steam reformer. Thus, the spent iron oxide catalyst is expected to contain only iron oxide, carbonaceous material, and catalyst supporting material. The estimated generation rate for this spent catalyst, based upon Exxon's estimate of the consumption rate (13), is 321 Mg/yr including 35 Mg/yr of inert balls.

Spent methanation catalyst from hydrogen generation (Stream 439) (Hydrogen production - Area 4)

The reaction of low concentrations of CO in a mixture with H_2 to form CH_4 is incorporated into the EDS base case design as a gas purification step for CO removal. The spent nickel oxide catalyst is expected to contain little or no contaminants because the gas is fairly clean by the time it enters the methanator. Also, carbon deposition is not known to be a problem in methanation. The estimated generation rate of this spent catalyst is 115 Mg/yr, including 18 Mg/yr inert balls (13).

Spent drying agents from ammonia synthesis (Stream 453) (Hydrogen production - Area 4)

The spent alumina used as drying agents for the feed gas to the ammonia synthesis reactor is generated only from the base case design. Although characterization data for this waste are not available, the spent alumina is not expected to contain any contaminants other than water and traces of ammonia. The estimated generation rate of the spent drying agents is 172 Mg/yr, including 27 Mg/yr of inert balls (13).

Spent ammonia synthesis catalyst (Stream 454) (Hydrogen production - Area 4)

Ammonia synthesis is used in the base case design to remove traces of nitrogen from the hydrogen produced. Typical catalysts used today for ammonia synthesis contain Al_2O_3, K_2O, CaO and MgO (total of 5-8%) as promoters in addition to iron and iron oxides (33). The spent catalyst generated is similar in composition to the fresh iron oxide catalyst, but is also expected to contain traces of NH_3 and ammonium compounds. Leaching characteristics for this spent catalyst are not known. The estimated generation rate of the spent ammonia synthesis catalyst is 166 Mg/yr, including 17 Mg/yr of inert balls.

Slag from the partial oxidation unit (Stream 442) (Hydrogen production - Area 4)

In the MFS case, about half of the ash in the coal fed to the EDS liquefaction section will appear as slag from the partial oxidation unit. The other half of the ash will be discharged in the various solid streams from the Flexicoking operations. Slag from the partial oxidation unit is provided to the slag handling system in two streams, a 90% solids coarse slag stream and a 20% solids fine slag slurry. After dewatering, coarse and fine slag fractions are combined to yield a 79% solids waste gasifier slag (Stream 442) to disposal. Characterization data are not available for slags from gasification of EDS vacuum-bottoms residues. The slag is anticipated to be similar in composition to the feed coal ash with up to 2% carbonaceous material. The estimated generation of the combined slag is 447,300 Mg/yr on 79% solids basis (13).

Leaching data have been published for slag produced in the Texaco pilot plant at Montebello, California. This slag resulted from gasification of SRC-II flash drum bottoms obtained with Kentucky No. 9/14 coal (27,34). Detailed leaching results are presented in Tables 3-58 and 3-59. Table 3-60 presents the content of selected polynuclear aromatic (PNA) hydrocarbons in the slag itself. Bulk and true specific gravities of the composite dry slag were found to be 1.64 and 2.62, respectively. Approximately 50% of the dry composite slag is smaller than 0.4 mm. Batch leaching tests were performed using slag/extractant weight ratios, time intervals, and equipment suggested in the proposed standard leaching tests (34). Although the two leaching tests with HNO_3 and NH_4OH are not standard, the same techniques and solid-to-liquid ratios were used. Column leaching tests were performed in 95 mm (3.75 inch) diameter columns using a slag depth of 305 mm (12 inches) and approximately one liter of demineralized water extractant. Results of these tests appear to indicate that the concentration levels of the pertinent trace elements in the leachate from the slag are well below RCRA EP criteria. Low levels of polynuclear aromatics were detected in the slag by benzene extraction, although details of the extraction procedure are not available.

Spent high temperature shift catalyst from hydrogen generation (Stream 444) (Hydrogen production - Area 4)

For the MFS case design, the synthesis gas fed to the shift converters enters the converters without acid gas removal, and thus still contains sour hydrogen sulfide (sour shift). The high temperature catalyst used for this application is a cobalt-molybdenum catalyst. The spent catalyst is similar

TABLE 3-58. ELEMENTAL ANALYSIS OF LEACHATES DERIVED FROM VARIOUS BATCH LEACHING PROTOCOLS APPLIED TO TEXACO SLAG FROM GASIFICATION OF SRC-II RESIDUE (Ky 9/14 Coal)

Element	EPA Extraction Procedure	ASTM H_2O	ASTM pH 5	0.1 N HNO_3	0.1 N NH_4OH	RCRA* Criteria
Arsenic, ppb	0.53	5	5	6.1	2.4	5000
Barium, ppm	0.5	0.5	0.5	0.5	0.5	100
Cadmium, ppb	3.4	4.0	4.9	25	1.3	1000
Chromium, ppb	0.13	3.7	3.9	42	1.9	5000
Lead, ppb	1.6	5.2	2.1	8.3	4.4	5000
Mercury, ppb	0.191	0.025	0.010	0.04	0.33	200
Nitrate (as N), ppm	---†	0.05‡	0.05‡	---	---	None
Selenium, ppb	1	0.04	0.04	1	7	1000
Silver, ppb	0.07	0.09	0.23	3.2	0.20	5000
Sulfate, ppm	4.89	23.63	22.90	26.79	36.57	None
Iron, ppm	20.5	147	720	2482	0.06	None
Copper, ppb	0.94	---	---	---	---	None
Beryllium, ppb	0.09	---	---	---	---	None
Antimony, ppb	2	---	---	---	---	None
Nickel, ppb	758	---	---	---	---	None

Data source: Table K.2 of Reference 34. ASTM H_2O uses distilled water, whereas ASTM pH 5 uses a sodium acetate -acetic acid buffer solution. The HNO_3 and NH_4OH tests are not standard tests.

* RCRA regulations (May 19, 1980).

† --- means not determined

‡ Samples not acidified before analysis.

**Extraction procedure published in Federal Register (December 18, 1978).

TABLE 3-59. ELEMENTAL ANALYSIS OF SLAG LEACHATES DERIVED FROM LABORATORY COLUMNS

Column Number	Element	Leach Number			
		1	2	10	20
1	Arsenic, ppb	1.16	0.79	2.0	3.1
2		1.36	1.33	1.53	3.4
1	Barium, ppb	<0.5	<0.5	<0.5	<0.5
2		<0.5	<0.5	<0.5	<0.5
1	Cadmium, ppb	11	0.13	0.014	0.10
2		16	0.11	0.020	0.091
1	Chromium, ppb	7.0	3.5	0.25	0.49
2		6.2	3.2	0.24	0.41
1	Lead, ppb	1.4	<0.3	<0.3	<0.43
2		0.93	<0.3	<0.3	<0.3
1	Mercury, ppb	0.007	0.024	0.013	0.019
2		0.011	0.012	0.003	0.001
1	Selenium, ppb	120	44	16	<5
2		195	34	<5	<5
1	Silver, ppb	0.34	0.17	0.13	0.10
2		0.36	0.17	0.12	0.10

Data source: Table K.3 of Reference 34. Data are from tests performed at the Texaco pilot plant at Montebello, California. The ash/slag was from gasification of a Kentucky No. 9/14 SRC-II flash drum residue. The laboratory columns used (3.75 in. diameter) have a 12 in. depth of slag and are dosed daily with 1 liter of demineralized water (about 1 column volume). Results for leaches 1, 2, 10 and 20 are given in this table.

TABLE 3-60. CONCENTRATIONS OF BENZENE SOLUBLES AND SELECTED POLYNUCLEAR AROMATICS IN TEXACO SLAG FROM GASIFICATION OF SRC-II RESIDUE

Contaminant	Coarse Slag	Fine Slag
Total benzene solubles, wt %	0.0315	0.0165
Benz-a-anthracene, ppb	12	10
Benz-a-pyrene, ppb	17	13
Benzo-e-pyrene, ppb	4	17
Benzo(g,h,i) perylene, ppb	8	5

Data source: Reference 34. Vacuum bottoms from the Fort Lewis SRC-II pilot plant were gasified in a Texaco gasifier. The gasifier slag was then subjected to a benzene extraction, and the benzene extract was analyzed for the presence of benzene solubles and polynuclear aromatic (PNA) compounds.

in composition to the fresh catalyst, except for the presence of contaminants such as carbonaceous material, sulfides, sulfur, and volatile trace metals. Leaching characteristics for this spent catalyst are not available. The estimated generation rate of the spent cobalt-molybdenum catalyst is 451 Mg/yr, including 35 Mg/yr of inert balls (13).

Spent low temperature shift catalyst from hydrogen generation (Stream 445) (Hydrogen production - Area 4)

The low temperature shift converter is employed in the MFS case design because considerable amount of carbon monoxide remains in equilibrium at the exit temperature of the preceding 2-stage high temperature shift converter. The low temperature shift catalyst used is also a cobalt-molybdenum catalysts. The spent catalyst is expected to contain cobalt, molybdenum, sulfides, carbonaceous material, and possibly volatile trace metals. Leaching characteristics for this spent catalyst are not known. The estimated generation rate of the spent Co/Mo catalyst is 254 Mg/yr, including 22 Mg/yr of inert balls (13).

3.3.5 Auxiliary Operations (Area 5)

The auxiliary operations in an EDS commercial plant are: 1) raw water treatment, 2) steam and power generation, 3) cooling operations, 4) oxygen production, and 5) product and by-product storage. These auxiliary operations and the associated waste streams are described in the ensuing subsections.

3.3.5.1 Raw Water Treatment (Auxiliary Operations - Area 5)

Various treatments are required to render raw waters to be suitable for use as boiler feed water, cooling tower makeup water and potable water in direct liquefaction facilities. The degree of treatment required depends upon the characteristics of the raw water and the end-use water quality requirements. Figure 3-8 presents a block flow diagram of the raw water treatment systems. Boiler feed water has highest quality requirements and thus requires more sophisticated treatment. In Table 3-61, the source and characteristic of raw waters assumed for the EDS direct liquefaction facility are presented.

Raw water is generally pumped from rivers and stored in a reservoir. This storage: 1) provides a reliable supply of water to the facility independent of river flow, 2) reduces the impact of raw water quality variation, and 3) allows sedimentation of silts and other large suspended materials.

The raw water is chlorinated to prevent biogrowth, to destroy organics, and to oxidize reduced species (mainly iron and manganese) to their more insoluble oxidized form for removal in the subsequent coagulation/sedimentation step. Alum and polymers are generally added in the coagulation step to improve suspended solids removal. After the coagulation/settling step, the water is generally suitable for use as cooling tower makeup. If high recirculation is required, acidification to reduce bicarbonates and/or addition of lime and magnesium hydroxide to remove hardness may be required.

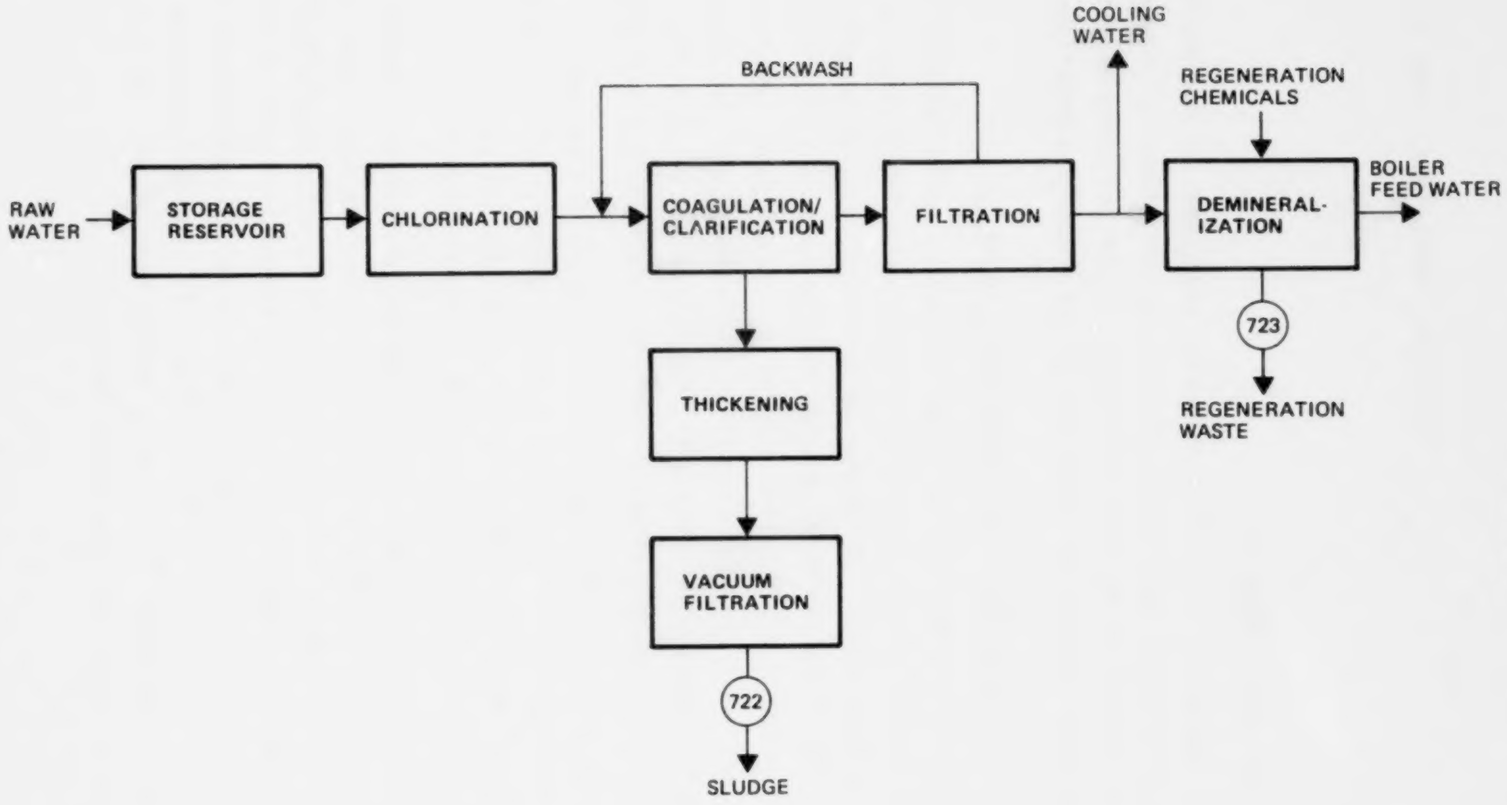
RAW WATER
STORAGE RESERVOIR
CHLORINATION
BACKWASH
COAGULATION/ CLARIFICATION
FILTRATION
COOLING WATER
REGENERATION CHEMICALS
DEMINERAL-IZATION
BOILER FEED WATER
723
REGENERATION WASTE
THICKENING
VACUUM FILTRATION
722
SLUDGE

Figure 3-8. Raw water treatment block flow diagram.

TABLE 3-61. ASSUMED CHARACTERISTICS OF RAW WATER FOR EDS DIRECT LIQUEFACTION FACILITY

Parameter*	Mississippi River
Total Hardness, as ppmw $CaCO_3$	130
M-Alkalinity	105
SO_4	46
Ca	37
Mg	9
Na	9
Cl	10
SiO_2	6
pH	7.7

* All concentrations except pH expressed in mg/l.

A filtration step is usually required to protect the demineralization units, and to render the water suitable for use as potable water. The filter effluent is sent to the demineralization unit. The demineralization effluent is essentially pure water and is used as boiler feed water.

3.3.5.1.1 Gaseous Waste Stream

There is no gaseous waste stream from raw water treatment.

3.3.5.1.2 Liquid Waste Stream

There is only one aqueous waste stream from raw water treatment:

- Stream 723 - regeneration waste from water demineralization.

Regeneration waste from water demineralization (Stream 723) (Raw water treatment - Area 5)

Mixed bed anion-cation exchangers are used for water demineralization in the EDS plant. In ion exchangers, dilute chemicals are introduced downward through the resin bed periodically to remove exchanged ions and to restore exchange capacity. The regeneration step results in the generation of a waste brine which must subsequently be handled as Stream 723. Based on the demineralization feedwater requirements (13), the estimated generation rate for the regeneration waste is 30,000 kg/hr (3.0 m^3/hr) for the base case design and 34,000 kg/hr (3.4 m^3/hr) for the MFS case design. The estimated characteristics of the regeneration waste are presented in Table 3-62. This waste stream contains high levels of dissolved solids.

3.3.5.1.3 Solid Waste Stream

There is only one solid waste stream from raw water treatment:

- Stream 722 - raw water treatment sludge.

Raw water treatment sludge (Stream 722) (Raw water treatment - Area 5)

The sludge from the filter press is a cake-like material with about 60% moisture content, and consists mainly of calcium carbonates and magnesium

TABLE 3-62. ESTIMATED CHARACTERISTICS OF REGENERATION WASTE FROM WATER DEMINERALIZATION (STREAM 723)

Contaminant	Concentration, ppmw
Calcium	1,030
Magnesium	760
Sodium	2,280
Sulfate	11,690
Chloride	530
Nitrate	100
Silica	70

Data source: Estimated using characteristics of raw water (Table 3-61).

hydroxide. The estimated characteristics of this stream for the EDS liquefaction facilities are presented in Table 3-63. Based on the raw water feed requirement (13), the estimated generation rate of the raw water treatment sludge is 46 Mg/day and 41 Mg/day for the base case and MFS case design, respectively.

3.3.5.2 Steam and Power Generation (Auxiliary Operations - Area 5)

The steam generation system provides net plant requirements and utilizes all the low Btu fuel gas available, which amounts to 1.46 TJ (1385 MM Btu) per hour for the base case design and 1.69 TJ (1599 MM Btu) per hour for the MFS case design. Additional coal feed, at the rate of 35.3 Mg/hr for the base case design and 92.8 Mg/hr for the MFS case design, are provided to the steam generation system as supplemental fuel. Steam is required in the EDS plant as feed to Flexicoking reactor and gasifier, as feed to the steam reformer (base case only) or partial oxidation unit (MFS case only), and for other process uses such as stripping steam for the atmospheric fractionator.

Although the EDS commercial design stipulated the purchase of power from a local utility, some coal liquefaction plants may prefer on-site power generation. For this reason, this PCTM assumes that power is generated through on-site coal-fired boilers. For the base case design, the power requirement of 223 MW can be met by the use of coal at a rate of 95.8 Mg/hr. For the MFS case design, the power requirement of 160 MW can be met by the use of coal at 68.7 Mg/hr. This power is required in the EDS plant to run electrically-driven equipment such as pumps and compressors.

If power were purchased from a local utility, instead of being generated on-site, the discharge streams discussed below -- associated with the coal-fired power boilers -- would be eliminated from the EDS power plant.

3.3.5.2.1 Gaseous Waste Streams

There are two gaseous waste streams from steam and power generation:

TABLE 3-63. ESTIMATED CHARACTERISTICS OF RAW WATER TREATMENT SLUDGE (STREAM 722)

Component	Wt % in Sludge
$CaCO_3$	36.4
$Mg\ (OH)_2$	3.6
H_2O	60

Data source: Estimated using characteristics of raw water (Table 3-61).

Auxiliary Operations - Area 5
Streams 701a, 707a

- Stream 701a - flue gas from steam generation system
- Stream 707a - flue gas from power generation system.

Flue gas from steam generation system (Stream 701a) (Steam and power generation - Area 5)

The burning of low-Btu fuel gas provides approximately 63% and 43% of the energy input for steam generation in the base case and MFS case designs, respectively. The remaining energy requirements for steam generation are provided by burning coal. Estimated emission rates of pollutants in the flue gas from the steam generation system are presented in Table 3-64. It is re-emphasized that these are emissions assuming no particulate or SO_2 controls. Emissions of particulate matter and SO_2 are mainly contributed by the coal feed. Assuming combustion with 30% excess air, the flow rate of the flue gas stream was calculated to be 42,560 kmol/hr (592,700 SCFM) for the base case design and 69,150 kmol/hr (962,930 SCFM) for the MFS case design. In these estimates, no assumptions were made concerning whether the plant steam needs will be generated by one large steam boiler, or two or more smaller ones.

Flue gas from power generation system (Stream 707a) (Steam and power generation - Area 5)

Estimated emission rates of pollutants in the flue gas from the power generation system are presented in Table 3-65, assuming no particulate or SO_2 controls. Emission rates of pollutants from the power generation system are higher than those from the steam generation system in the base case design, but lower in the MFS case design. Assuming combustion with 30% excess air, the flow rate of the flue gas stream was calculated to be 36,450 kmol/hr (507,600 SCFM) for the base case design, and 26,110 kmol/hr (363,600 SCFM) for the MFS case design.

3.3.5.2.2 Liquid Waste Streams

There are two aqueous waste streams from steam and boiler generation:

TABLE 3-64. UNCONTROLLED FLUE GAS EMISSIONS FROM STEAM GENERATION SYSTEM (STREAM 701a)

Pollutant	Emissions Rate, kg/hr	
	Base Case	MFS Case
Particulate Matter	2,190	5,740
NO_x (as NO_2)	297-366	670-750
CO	28-34	59-65
Hydrocarbons	7.2	16-1
SO_2	2,520	6,530

Data source: Emission rates for particulate matter were computed based on the ash content of coal and assuming that 81.3% of the coal ash results in fly ash; emission rates for SO_2 were computed based on the sulfur content of coal and assuming that 94% of the fuel sulfur becomes SO_2; emission rates for hydrocarbons, and CO were computed using AP-42 emission factors for pulverized coal-fired boilers and natural gas-fired boilers larger than 100 MM Btu/hr in capacity (21). Emission rates for NO_x from low-Btu gas combustion were computed by using AP-42 NO_x emission factor for natural gas-fired boilers larger than 100 MM Btu/hr in capacity (21). The NO_x emission factor used is 52-99 ng/J (0.12 - 0.23 lb/MM Btu), which is approximately the same as the 86 ng/J (0.20 lb/MM Btu) NSPS NO_x standard for utility boilers burning gaseous fuel. Emission rates for NO_x from coal combustion were computed using the NSPS NO_x standard of 260 ng/J (0.60 lb/MM Btu) for utility boilers burning bituminous coal.

Other assumptions: For the base case design, approximately 63% of the energy input to the steam boiler is provided by burning low-Btu fuel gas (125 Btu/SCF, 83 ppmv S) and 37.0% provided by burning coal (3.7% S, 7.6% ash). For the MFS case design, approximately 42.8% of the enrgy input to the steam boiler is provided by burning low-Btu fuel gas, and 57.2% provided by burning coal.

TABLE 3-65. UNCONTROLLED FLUE GAS EMISSIONS FROM POWER GENERATION SYSTEM (STREAM 707a)

Pollutant	Emission Rate, kg/hr	
	Base Case	MFS Case
Particulate Matter	5,930	4,250
NO_x (as NO_2)	601	431
CO	47.9	34.4
Hydrocarbons	14.3	10.3
SO_2	6,660	4,780

Data source: Emission rates for particulate matter were computed based on the ash content of coal and assuming that 81.3% of the coal ash results in fly ash; emission rates for SO_2 were computed based on the sulfur content of coal and assuming that 94% of the fuel sulfur becomes SO_2; emission rates for hydrocarbons, and CO were computed using AP-42 emission factors for pulverized coal-fired boilers larger than 100 MM Btu/hr in capacity (21). Emission rates for NO_x were computed using the NSPS NO_x standard of 260 ng/J (0.60 lb/MM Btu) for utility boilers burning bituminous coal. All power is assumed to be generated by burning coal containing 3.7% sulfur and 7.6% ash.

- Stream 702 - blowdown from steam generation system
- Stream 708 - blowdown from power generation system.

Blowdown from steam generation system (Stream 702) (Steam and power generation - Area 5)

Based on Exxon's design estimate (13), blowdown from the steam generation system is generated at the rate of 67 m^3/hr for the base case design and 55 m^3/hr for the MFS case design. This is a stream containing typically up to 4,000 mg/l total dissolved solids (35), and additives for scale and corrosion control such as hydrazine and phosphates.

Blowdown from power generation system (Stream 708) (Steam and power generation - Area 5)

Based on a typical boiler blowdown rate of 12.6 l/MW-hr (36), Stream 708 was estimated to be generated at the rate of 2.8 m^3/hr for the base case design and 2.0 m^3/hr for the MFS case design. Because higher pressure steam is generated in the power boilers, the blowdown contains lower levels of total dissolved solids in the 200-500 mg/l range (35). Additives for scale and corrosion control such as hydrazine and phosphates are also present.

3.3.5.2.3 Solid Waste Streams

There are two solid waste streams from steam and power generation:

- Stream 704 - bottom ash from steam generation system
- Stream 710 - bottom ash from power generation system.

Bottom ash from steam generation system (Stream 704) (Steam and power generation - Area 5)

Assuming that 18.7% of the ash content of the coal fed to the steam generation system results in bottom ash (36), the generation rate of this solid waste was calculated to be 3,960 Mg/yr for the base case design and 10,400 Mg/yr for the MFS case design. Concentrations of trace elements present in the bottom ash are similar to those found in the coal feed (Table 2-2), although the levels of some of the more volatile elements (or elements associated with volatile compounds) will be depleted (36).

Bottom ash from power generation system (Stream 710) (Steam and power generation - Area 5)

Using the same assumptions as for the steam generation system, the generation rate of bottom ash from the power generation system was calculated to be 10,740 Mg/yr for the base case design and 7,700 Mg/yr for the MFS case design.

3.3.5.3 Cooling Operation (Auxiliary Operations - Area 5)

All direct liquefaction plants generate non-recoverable heat, most of which is dissipated through cooling towers. The selection/design of optimum cooling systems would require detailed heat balances for the whole plant, and is beyond the scope of this study. The use of wet cooling towers is proposed in the Exxon design; this is the only type of cooling operation analyzed in the PCTM. The total plant wet cooling loads for the EDS commercial plant amount to a cooling water circulation rate of 55,190 m^3/hr for the base case design and 44,740 m^3/hr for the MFS case design.

A wet cooling tower removes heat by evaporating part of the circulating water. To prevent excessive buildup of impurities in the circulating circuit, some of the circulating water must be removed from the system. Makeup water is required to compensate for losses through evaporation, blowdown, and to a lesser extent, drift. The makeup water can either be from treated raw water or from treated process wastewater. Thus, cooling operation not only affects the overall plant water balance, but can also be a critical factor in the disposal and reuse of process wastewater.

3.3.5.3.1 Gaseous Waste Stream

There is only one gaseous waste stream from cooling operation:

- Stream 731 - drift and evaporation from cooling tower.

Drift and evaporation from cooling tower (Stream 731) (Cooling operation - Area 5)

Assuming that approximately 1.25 percent of the circulating cooling water is lost due to drift and evaporation, the total amount of water loss through this waste stream was calculated to be 690 m^3/hr for the base case design and 560 m^3/hr for the MFS case design. Only a small fraction of the cooling tower loss is due to drift alone. In modern cooling towers, drift losses are typically only 0.005 percent of the circulating water (35). Thus, drift losses only amount to 2.8 m^3/hr and 2.2 m^3/hr of water for the base case and MFS case designs, respectively; the remainder of the total loss is due to evaporation. Estimated characteristics for the drift droplets are presented in Table 3-66. There is also some potential for VOC emissions from cooling towers, because organics could be present in the cooling loop via heat exchanger leaks, etc.

3.3.5.3.2 Liquid Waste Stream

There is only one aqueous waste stream from cooling operation:

- Stream 732 - cooling tower blowdown.

Cooling tower blowdown (Stream 732) (Cooling operation - Area 5)

Blowdown from the cooling tower is necessary to maintain the total dissolved solids (TDS) in the circulating water at a reasonable level to prevent solids deposition. For estimation purposes, the TDS level in the circulating cooling water is assumed to be maintained at 500 mg/l, which corresponds to two cycles of concentration. The calculated cooling tower blowdown rate is 630 m^3/hr for the base case design, and 510 m^3/hr for the MFS case design. Estimated characteristics of the cooling tower blowdown are presented in Table 3-66.

3.3.5.3.3 Solid Waste Stream

There is no solid waste stream from cooling operation.

TABLE 3-66. CHARACTERISTICS OF COOLING TOWER DRIFT AND BLOWDOWN (STREAMS 731/732)*

Parameter	Concentration, mg/l
Hardness, as $CaCO_3$	260
TDS	500
Alkalinity, as $CaCO_3$	205
Cr	0.8
Cu	0.25
Fe	0.8
Mg	18
Na	18
Ni	0.03
Si	9
Zn	1
Sulfate	300
Chloride	20
Nitrate	8
Phosphate-P	4

Data source: Estimated from raw water characteristics by assuming two cycles of concentration, except that data for Cr, Cu, Fe, Ni, Zn, nitrate and phosphate were obtained from Reference 36.

* Concentration values only apply to the drift portion of Stream 731. The evaporation portion of Stream 731 is essentially free of contaminants, unless process wastewaters are used as makeup or organics have worked their way into the cooling loop via heat exchanger leaks, etc. in which case some volatilized organics might appear.

3.3.5.4 Oxygen Production (Auxiliary Operations - Area 5)

Oxygen production only applies to the MFS case design. Oxygen required for partial oxidation is assumed to be produced by standard cryogenic air separation units. Air is compressed to 0.58 to 0.61 MPa (85-90 psia) and cryogenically cooled to facilitate distillation of oxygen, nitrogen and noble gases. The oxygen stream, containing small quantities of nitrogen and argon, is compressed and sent to the gasifiers. Air and oxygen compressors can either be steam, gas, electric driven or a combination thereof. The separated nitrogen containing small quantities of oxygen, water and carbon dioxide is primarily vented to the atmosphere. However, a portion of the nitrogen stream may be utilized as an inerting agent for coal storage and transfer, and/or as stripping gas for solvent regeneration in acid gas treatment and phenol recovery units. The quantity of condensate resulting from air compression depends upon atmospheric humidity and therefore, is highly variable. Condensate contains only dissolved gases and can be utilized as a supplement to the plant high quality water supply. The estimated EDS MFS case oxygen requirement is 4,190 Mg/day (13).

3.3.5.4.1 Gaseous Waste Stream

A gaseous stream containing mostly nitrogen is generated. As discussed previously, this stream may be used as an inerting agent in other plant areas or may be vented to the atmosphere and is not considered a waste stream. There is no gaseous waste stream from oxygen production.

3.3.5.4.2 Liquid Waste Stream

The only liquid stream generated is a condensate which may be used as makeup water in other plant areas. There is no liquid waste stream from oxygen production.

3.3.5.4.3 Solid Waste Stream

Dessicants are used in oxygen production to remove VOCs from incoming air.

Also, particulate filters are used. Thus, small amounts of solid wastes will be generated from discarded dessicants and particulate filter material on an intermittent basis.

3.3.5.5 Product and By-Product Storage (Auxiliary Operations - Area 5)

Storage capacities for product and by-products were estimated based upon the expected production rates of upgraded liquid and gaseous fuels for the EDS direct liquefaction plant. Storage for the EDS process plant consists of a variety of storage tanks for liquid products, by-products, and makeup chemicals. Depending on the vapor pressure of the liquid stored, the design incorporates floating roof, fixed (cone) roof, pressurized or dome type tanks. The more volatile products (e.g., LPG, ammonia) are stored in pressure vessels and are expected to have no routine evaporative emissions. Products such as naphtha are stored in floating roof tanks and products such as phenol and blended fuel oil are stored in fixed roof tanks. Some streams such as sour water/ phenolic water require intermediate storage and are stored in dome tanks. Various products such as vacuum bottoms, liquefaction hydrogenated recycle solvents, blended fuel oil, etc., require storage at temperatures ranging from 343 to 533°K (160°F to 500°F). Consequently, these tanks are provided with a heating mechanism and are insulated in designs proposed by Exxon.

3.3.5.5.1 Gaseous Waste Stream

There is only one type of gaseous waste stream from product and by-product storage:

- Stream 751 - evaporative emissions from product and by-product storage.

Evaporative emissions from product and by-product storage (Stream 751) (Product and by-product storage - Area 5)

For both the base case and MFS case designs, storage tanks containing fuel oil, naphtha, and phenol were identified as the main sources of evaporative emissions. Based on Exxon's proposed design, naphtha will be stored in external floating roof tanks, and blended fuel oil and phenol will be stored in fixed roof tanks.

Evaporative emissions from external floating roof tanks consist primarily of standing storage losses and withdrawal losses. Standing storage losses result from wind induced mechanisms as air flow across the top of an external floating roof tank. Withdrawal losses are caused by the vaporization of liquid that clings to the tank wall and is exposed to the atmosphere when a floating roof is lowered by withdrawal of liquid. Evaporative emissions from fixed roof tanks consist primarily of breathing losses and working losses. Breathing losses are caused by the expulsion of vapor from a tank due to vapor expansion and contraction from changes in temperature and barometric pressure. Working losses are the combined losses from tank filling and emptying.

Estimated uncontrolled evaporative emissions from product and by-product storage are presented in Table 3-67. In estimating these uncontrolled emissions, the external floating roof tanks were assumed to be equipped with primary seals only, and the fixed roof tanks were assumed to be neither equipped with internal floating roofs nor with vapor recovery systems.

Data on the components of evaporative emissions associated with storage of coal derived liquid fuels are generally lacking. However, limited data are available regarding the chemical composition of some of the direct liquefaction products. For example, the major constituents of naphtha are paraffins, naphthenes and aromatics, and these are also expected to be present in evaporative emissions. Compositions of these products are presented in Section 3.3.6.

3.3.5.5.2 Liquid Waste Stream

There is no liquid waste stream from product and by-product storage.

3.3.5.5.3 Solid Waste Stream

Tanks will require cleaning at one time or another to remove sludges from storage of hydrocarbon products. Thus, small amounts of waste sludges will be generated on an intermittent basis.

TABLE 3-67. UNCONTROLLED EVAPORATIVE EMISSIONS FROM STORAGE TANKS (STREAM 751)

Product	Capacity, m^3/tank	No. of Tanks	Type of Tanks	Total Uncontrolled Emissions* kg/hr
Base Case				
Naphtha	41,200	2	Floating roof	2.14
Blended Fuel Oil	36,800	2	Fixed roof	0.018
Phenol	730	2	Fixed roof	0.022
MFS Case				
Naphtha	40,000	2	Floating roof	2.14
Blended Fuel Oil	34,000	2	Fixed roof	0.016
Phenol	730	2	Fixed roof	0.022

* Uncontrolled emissions from fixed roof and floating roof storage tanks were calculated based on emission equations listed in AP-42 (21).

3.3.6 Products

Liquid products from the EDS plant include C_3 LPG, C_4 LPG, naphtha, and blended fuel oil. The C_3 LPG, C_4 LPG, and blended fuel oil are produced as finished products at the coal liquefaction plant according to either the base case or the MFS case design. The naphtha product, however, requires further processing prior to use as gasoline blendstock or in aromatics production. In addition, a pipeline gas product is generated from the MFS case design. The product yields, origins, and composition information are presented in this section. By-products from pollution control operations include sulfur, ammonia and crude phenol. These by-products are not covered here, but are briefly addressed in Section 4 where the responsible control operations are discussed.

3.3.6.1 Pipeline Gas

Pipeline gas is obtained from the hydrogen recovery section (Hydrogen plant - Area 4) of the MFS case design. The hydrogen recovery section cryogenically separates hydrogen, C_1/C_2 pipeline gas product, and C_3^+ product gas from the high pressure purge gases and low-pressure off-gases from liquefaction, solvent hydrogenation, and Flexicoking. The pipeline gas becomes a net product in the MFS case design because the C_1/C_2 gas is not needed for hydrogen production by steam reforming, as in the base case design.

Total pipeline gas product from the MFS case design amounts to 1,991,000 Nm^3/stream day (70.3 MM SCFD, or 11,900 BFOE*). The pipeline gas contains 63.0 vol. % methane, 22.3 vol. % C_2, 2.7 vol % C_3^+ hydrocarbons, 5.4 vol. % N_2, 3.4 vol. % H_2, and 3.2 vol. % CO, and has a lower heating value (LHV) of 40.3 MJ/Nm^3 (1024 Btu/SCF) (13). Thus, this pipeline gas is similar to natural gas in that methane is the primary constituent. The EDS pipeline gas, however, contains more C_2 and C_3 hydrocarbons and CO than natural gas.

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* 1 barrel fuel oil equivalent (BFOE) = 6,050,000 Btu (LHV)

3.3.6.2 LPG

Two LPG products, C_3 LPG and C_4 LPG, are produced from the product recovery section (Product separation and purification - Area 3) of the EDS commercial plant. The C_3 LPG contains 95.8 wt % C_3, 0.9 wt % C_2^-, and 3.3 wt % C_4^+ (13). The C_4 LPG contains 95.1 wt % C_4, 3.2 wt % C_3^-, and 1.7 wt % C_5^+(13). For the base case design, the production rates of C_3 LPG and C_4 LPG are 0.236 Gg (463 m^3; 2,915 bbl or 1,780 BFOE) per stream day and 0.227 Gg (393 m^3; 2,465 bbl or 1,680 BFOE) per stream day, respectively (13). For the MFS case design, the production rates of C_3 LPG and C_4 LPG are 0.334 Gg (667 m^3; 4,198 bbl or 2,570 BFOE) per stream day and 0.235 Gg (405 m^3; 2,548 bbl or 1,740 BFOE) per stream day, respectively (13).

3.3.6.3 Naphtha

Naphtha is obtained from the product recovery section (Product separation and purification - Area 3) of the EDS commercial plant. In product recovery, the total distillate products from the atmospheric fractionator, the solvent stripper, and the Flexicoker fractionator are fed to the deethanizer. The bottoms product from the deethanizer is then fed to the debutanizer to yield a stabilized naphtha product.

The naphtha product is estimated by Exxon to contain mainly the pentanes to 350°F fraction. Elemental composition of this product is as follows: 85.20 wt % C, 13.16 wt % H, 0.43 wt % S, 0.06 wt % N, and 1.15 wt % O (13). Naphtha was one of the two product streams sampled and analyzed by EPA at the EDS pilot plant (37). However, the naphtha sampled was a distillate cut from the solvent fractionation tower and does not necessarily represent the naphtha produced from other process operations. In particular, the aromatic content of this naphtha sample might be higher than in naphtha from commercial EDS facilities.

In Table 3-68, concentrations of trace metals present in this naphtha sample are presented. In Table 3-69, results of organic analysis for the naphtha sample are presented. These results indicated that approximately 80 percent of the total organic concentration of the naphtha was aliphatic hydrocarbons. The most prevalent aromatic compounds were benzenes and substituted benzenes. Additional analyses for volatile organics also showed that about 80 percent of the total organic concentration of the naphtha was volatile organics, as would be expected for this light oil (37).

Based on Exxon's estimate (13), naphtha is produced at the rate of 2,880 Gg (3,770 m^3; 23,710 bbl) per stream day for the base case design, and 2,824 Gg (3,695 m^3; 23,243 bbl) per stream day for the MFS case design.

3.3.6.4 Blended Fuel Oil

The low sulfur fuel oil (LSFO) from the EDS commercial plant is obtained by blending: 1) vacuum gas oil from the vacuum fractionation tower (Product separation and purification - Area 2); 2) LSFO from the solvent fractionation tower (Product separation and purification - Area 2); and 3) LSFO from Flexicoking (Liquefaction residue processing/hydrogen production - Area 4). Vacuum gas oil may also be further processed by hydrocracking or catalytic cracking to produce lighter products, but additional processing is not considered in the current designs.

The fuel oil products from the base case and MFS case designs are slightly different. In Table 3-70, the compositions of fuel oil products from the two designs are compared. The fuel oil from the base case design contains a higher percentage of the 1000°F^+ fraction, and also slightly higher levels of sulfur and nitrogen.

The LSFO from the solvent fractionation tower was the other product stream sampled and analyzed by EPA at the EDS pilot plant (37). As in the case of naphtha, the light solvent fuel oil sampled only represents part of the blended

TABLE 3-68. CONCENTRATIONS OF TRACE METALS IN NAPHTHA FROM EDS PILOT PLANT SOLVENT FRACTIONATION TOWER

Element	Concentration, mg/l
Barium	0.013
Calcium	0.67
Iron	4.0
Magnesium	0.16
Manganese	0.029
Nickel	0.049
Potassium	0.32
Sodium	0.72

Data source: Reference 37.

TABLE 3-69. CONCENTRATIONS OF ORGANIC COMPOUNDS IN NAPHTHA FROM EDS PILOT PLANT SOLVENT FRACTIONATOR TOWER

Organic Compound Class	Concentration, g/l
Aliphatic hydrocarbons	907.96
Ethers	9.55
Aldehydes and Ketones	2.88
Benzenes and Substituted Benzene Hydrocarbons	127.4
Phenols*	8.1
Heterocyclic Oxygen Compounds	47.8
Heterocyclic Sulfur Compounds (Thiophenes)	18.05
Total	1,121.74

Data source: Reference 37.

* Breakdown for the phenols: 1.3 g/l o-cresol, 5.9 g/l C_2 phenol, and 0.9 g/l C_3 phenol.

Note: Aromatic content of this naphtha sample might be higher than in naphtha from commercial EDS facilities.

TABLE 3-70. COMPARISON OF COMPOSITION OF FUEL OIL PRODUCTS FROM THE EDS BASE CASE AND MFS CASE DESIGNS

Component	Base Case	MFS Case
Composition		
C_5/400°F, wt %	19.0	19.2
400/700°F, wt %	34.0	35.2
700/1000°F, wt %	29.1	31.6
1000°F, wt %	17.9	14.0
Elemental Composition		
C, wt %	87.56	87.72
H, wt %	8.83	8.89
S, wt %	0.54	0.51
N, wt %	0.77	0.75
O, wt %	2.30	2.13

Data source: Tables 1-S-2 and 2-S-2 of Reference 13.

fuel oil mix. In Table 3-71, concentrations of trace metals present in this light solvent fuel oil sample are presented. In Table 3-72, preliminary results of organic analysis for the light solvent fuel oil sample are presented. Although complete analysis results are not yet available, the results to-date indicated that the major organic components are: aliphatic hydrocarbons, benzene and substituted benzene hydrocarbons consisting mostly of tetralin and methyl tetralin, phenols, and heterocyclic oxygen compounds such as furans and especially C_1-C_5 benzofurans (13). Comparable data for the vacuum gas oil (VGO) and LSFO from Flexicoking are not publicly available. However, it is anticipated that the blended fuel oil will contain more of the heavier hydrocarbon fractions because both the VGO and the LSFO from Flexicoking are heavier (13).

Based on Exxon's estimate (13), blended fuel oil is produced at the rate of 6.033 Gg (5,868 m^3; 36,910 bbl or 36,680 BFOE) per stream day for the base case design, and 5.788 Gg (5,625 m^3; 35,378 bbl or 35,160 BFOE) per stream day for the MFS case design.

TABLE 3-71. CONCENTRATIONS OF TRACE METALS IN LIGHT SOLVENT FUEL OIL FROM EDS PILOT PLANT SOLVENT FRACTIONATION TOWER

Element	Concentration, mg/l
Aluminum	0.41
Barium	0.018
Boron	0.18
Calcium	1.2
Chromium	0.12
Copper	0.032
Iron	1.0
Magnesium	0.16
Manganese	0.028
Nickel	0.092
Potassium	0.36
Sodium	1.0
Strontium	0.006
Titanium	0.016

Data source: Reference 37.

TABLE 3-72. CONCENTRATIONS OF ORGANIC COMPOUNDS IN LIGHT SOLVENT FUEL OIL FROM EDS PILOT PLANT SOLVENT FRACTIONATION TOWER

Organic Compound Class	Concentration, g/l
Aliphatic Hydrocarbons	54.64
Ethers	0.051
Aldehydes and Ketones	0.026
Thiols, Sulfides and Disulfides	0.002
Benzene and Substituted Benzene Hydrocarbons	58.74
Phenols*	19.06
Fused Polycyclic Hydrocarbons†	11
Fused Non-Alternant Polycyclic Hydrocarbons ‡	1.4
Heterocyclic Nitrogen Compounds	0.0091
Heterocyclic Oxygen Compounds	140.21
Heterocyclic Sulfur Compounds (Thiophenes)	0.023
Total	285.16

Data source: Reference 37. These results are preliminary.

* Breakdown for the phenols: 0.059 g/l phenol, 7.7 g/l methyl naphthol, 5.5 g/l C_2 naphthol, 5.4 g/l C_3 naphthol, and 0.4 g/l C_4 naphthol.

† The only fused polycyclic hydrocarbon identified was methyl naphthalene.

‡ The only fused non-alternant polycyclic hydrocarbon identified was fluorene.

3.3.7 Fugitive Organic Emissions

Fugitive organic emissions from the EDS commercial plant are not derived from any one process area alone. For this reason, these emissions are discussed in this separate section.

3.3.7.1 Air Fugitives

There are many potential sources of fugitive organic air emissions in a typical EDS direct coal liquefaction plant. Some of these are: pumps, compressors, in-line process valves, pressure relief devices, open-ended valves, sampling connections, and flanges. Extensive tests and measurements for fugitives have been performed at petroleum refineries (38). As a result of this testing, average emission factors were developed for fugitive emission sources such as pump seals, compressor seals, valves, etc. (39). These factors were applied to synfuel plants because it was assumed that synfuel plant operations are similar to petroleum refining operations. However, it should be recognized that emissions of fugitive organics from EDS plants and petroleum refineries could differ in two aspects: 1) emission factors could differ since coal liquefaction slurries are abrasive and could cause more leaks in some parts of the plant than conventional refinery experience would suggest; and 2) compositions of fugitive organic emissions could differ significantly, depending on the compositions of process streams. Limited data are available on the concentrations of organic compounds in some of the wastewater streams (e.g., Table 3-14 for liquefaction cold separator wastewater) and product streams (e.g., Table 3-69 for naphtha) from the 250 ton coal/day EDS pilot plant. Fugitive organic emissions from the EDS commercial plant are expected to contain some of the same constituents as found in the wastewater streams and product streams, such as aliphatic hydrocarbons, benzene and substituted benzene hydrocarbons, and phenols.

Fugitive emissions estimates were made by estimating the number of each type of emission source and applying an emission factor with no adjustment for size, pressure or flow rate. Estimates of the number of valves, flanges, pump seals, compressor seals, drains, relief valves and sampling connections in the EDS commercial plant were provided by Exxon (40).

The process streams associated with each piece of equipment were classified with respect to percent hydrocarbon content and hydrocarbon type. Different factors were used for liquid streams containing light and heavy hydrocarbons. Liquid streams containing C_2 through C_9 hydrocarbons, naphtha and other light aromatic hydrocarbons were classified as light. Coal liquids, low sulfur fuel oil, and other heavy hydrocarbons were classified as heavy. All emissions factors shown assume 100% hydrocarbon content, so all emission factors except those for compressors were multiplied by the actual hydrocarbon content for each process stream. Streams containing less than 10% hydrocarbons were neglected and those containing greater than 80% hydrocarbons were considered to contain 100% hydrocarbons. Gaseous streams were classified as either hydrocarbon or hydrogen depending on which was present in a greater percentage. The compressor seal emission factors for these two classifications were used as given and not altered by percent hydrocarbon content.

Results of these fugitive emission calculations for eht EDS process are given in Table 3-73. Since separate component counts distinguishing the two EDS designs are not available, it is assumed that the total fugitive hydrocarbon emissions of 153 kg/hr apply equally to the base case design or the MFS case design. Also, only a small portion of the total uncontrolled hydrocarbon emissions in Table 3-73 are likely to be low-volatile heavy organics, which would be expected to contain most of the compounds of health concern.

3.3.7.2 Liquid Fugitives

Liquid fugitives from the EDS commercial plant include spills, leaks, and area runoffs. At the present time, it is not possible to estimate the

TABLE 3-73. UNCONTROLLED FUGITIVE HYDROCARBON EMISSIONS FROM EDS COMMERCIAL PLANT

Components/ Service	Total No. of Components	Emission Factor, * kg/hr/unit	Uncontrolled Hydrocarbon Emissions kg/hr
Pump Seals			
Light Liquid	78	0.113	8.85
Heavy Liquid	164	0.0209	3.40
Compressor Seals			
Hydrocarbon	57	0.635	36.2
Hydrogen	18	0.0499	0.90
Drains			
Light Liquid	150	0.0318	4.76
Heavy Liquid	300	0.0454	13.61
Relief Valves	252	0.0862	21.73
Flanges	25,535	0.000254	6.49
Valves			
Gas/Vapor	1,211	0.0268	32.39
Light Liquid	1,787	0.0109	19.46
Heavy Liquid	3,575	0.000227	0.82
Hydrogen	519	0.00816	4.22
Sampling Connections	141	0.00227	0.32
Total			153.15

Data source: Component counts of emission sources are provided by Exxon (40); emission factors are obtained from Reference 39.

* These emission factors were determined using test data from a statistically valid sample of petroleum refineries, and reflect the average of a range of equipment types and maintenance practice.

nature of liquid fugitives caused by leaks or spills. Based on the maximum 25 year, 24 hour rainfall of 5.1 inches and 100% runoff, Exxon estimated a maximum dirty rain runoff (low quality rain runoff - Stream 901) of 318 m^3/hr during the period of the rainfall. This low quality rain runoff is expected to contain components present in the process and product/by-product streams, oil and grease, and moderately high levels of total dissolved solids.

3.3.8 Other Areas

There are a number of other operations in the EDS commercial plant that generate waste streams, especially in the pollution control operations process area. Waste streams from these operations are considered secondary discharge streams and are not covered here because Section 3 only addresses uncontrolled waste streams.

Some of the major pollution control operations that generate waste streams are:

- Sulfur recovery plant and associated tail gas treatment plant
- Sour water stripping/ammonia recovery system
- Phenolic extraction unit
- Wastewater treatment operations
- Particulate control and flue gas desulfurization systems for steam and power generation
- Incinerators/flaring for organic and CO containing gases
- Incinerators for solid residue/sludge disposal
- Solid waste disposal area.

Waste streams from these operations are discussed in Section 4 of this document.

3.4 SUMMARY OF UNCONTROLLED DISCHARGES BY MEDIUM

The uncontrolled discharges from the EDS commercial plant are summarized in tabular form in this section. The summary provided is broken down by medium and by process area. All uncontrolled waste streams, including intermittent discharges, are listed. In the summary tables, estimated uncontrolled mass emission rates of selected pollutants are presented. These pollutants were selected for presentation on the basis that they are generally of interest in the selection and evaluation of pollution control technologies. Detailed characterization of the uncontrolled waste streams which includes additional information on individual compounds has been presented in Section 3.3. Secondary waste streams from control technologies are discussed in Section 4 and not addressed here. The emission rates presented correspond to an EDS commercial plant processing 1,134 Mg/hr (30,000 tons per stream day) of "as received" Illinois No. 6 coal in the liquefaction area. Discharges shown are uncontrolled in order to facilitate consideration of control alternatives in Section 4. Of course, in a commercial EDS plant, suitable controls will be placed on these streams.

3.4.1 Gaseous Waste Streams

The total molar flow rate and mass emission rates for gaseous waste streams are summarized in this subsection according to process area. Mass emission rates for nine pollutants are presented: total hydrocarbons, CO, H_2S, NH_3, COS, SO_2, HCN, NO_x (as NO_2), and particulate matter.

3.4.1.1 Base Case

The characteristics of uncontrolled gaseous waste streams for the EDS base case design are summarized in Table 3-74. To avoid duplicate accounting, the Flexicoking gasifier/heater sour fuel gas (Stream 304) is assumed to be untreated and contains all the sulfur present as H_2S and COS, whereas flue gas streams resulting from the combustion of this low-Btu fuel gas are assumed to contain levels of SO_2 corresponding to the treated fuel gas (80 ppmv COS and 3 ppmv H_2S). Also, two totals are shown in Table 3-74, one for normal operation (no intermittent streams), another for the hypothetical worst case if all intermittent streams are discharging at the same time.

TABLE 3-74. SUMMARY OF UNCONTROLLED GASEOUS WASTE STREAMS FROM EDS COMMERCIAL PLANT (ILLINOIS COAL BASE CASE)

Uncontrolled Waste Stream	Total Flow, kmol/hr	Emission Rate, kg/hr								
		Total Hydrocarbon	CO	H_2S	NH_3	COS	SO_2	HCN	NO_x	Particulate Matter
Coal Preparation Process Area (Area 1)										
Stream 011 - fugitive dust from coal pile	N/A*	0	0	0	0	0	0	0	0	51
Stream 013 - fugitive particulates from coal handling and crushing	N/A	0	0	0	0	0	0	0	0	161-2,548
Coal Liquefaction Process Area (Area 2)										
Stream 102 - slurry drier vent gas	532	2,129	0	0.54	0	0	0	0	0	0
Stream 107 - flue gas from liquefaction slurry preheat furnace	34,535	2.4	14-20	0	0	0	91	0	97-186	4-12
Stream 803 - transient waste gas from liquefaction reactor (intermittent)**	3,280	30,836	426	2,849	2.7	0	0	0	0	0
Product Separation and Purification Process Area (Area 3)										
Stream 153 - vacuum fractionator off-gas†††	69	1,512	0	117	0	0	0	0	0	0
Stream 203 - flue gas from solvent hydrogenerator fuel preheat furnaces	3,864	0.27	1.5-2.3	0	0	0	10	0	11-21	0.45-1.4
Stream 508 - acid gas from DEA regenerator	1,214	0	0	26,562	812	27	0	0	0	0
Liquefaction Residue Processing/Hydrogen Production Process Area (Area 4)										
Stream 304 - Flexicoking gasifier/heater sour fuel gas †††	54,781	13,336	275,509	7,811	0	354	0	0	0	0
Stream 426 - vent gas from CO_2 removal by Catacarb process	6,100	35	27	0	0	0	0	0	0	0
Stream 434 - flue gas from hydrogen plant reformer furnaces	58,022	4.1	23-34	0	0	0	150	0	163-313	7-21
Stream 438 - hydrogen plant deaerator vent	6.5	0	<0.5	0	0	0	0	0	0	0
Stream 446 - regeneration/decommissioning off-gas from reformer catalyst (intermittent)††	6,800	No data‡	No data	0	0	0	0	0	No data	No data
Stream 448 - decommissioning off-gas from methanation catalyst (intermittent)‡‡	5,000	No data	No data	0	0	0	0	0	No data	No data
Stream 801 - transient waste gas from Flexicoking (intermittent)***	671	5,387	748	239	0	0	0	0	0	0

(Continued)

TABLE 3-74. (Continued)

Uncontrolled Waste Stream	Total Flow, kmol/hr	Emission Rate, kg/hr								
		Total Hydrocarbon	CO	H_2S	NH_3	COS	SO_2	HCN	NO_x	Particulate Matter
Auxiliary Operations Process Area (Area 5)										
Stream 701a - flue gas from steam generation system	42,560	7.2	28 - 34	0	0	0	2,522	0	297 - 366	2,187
Stream 707a - flue gas from power generation system	36,450	14.3	48	0	0	0	6,659	0	601	5,927
Stream 731 - drift and evaporation from cooling tower	N/A	No data	0	0	0	0	0	0	0	No data
Stream 751 - evaporative emissions from product/ byproduct storage	0.02†	2.2	0	0	0	0	0	0	0	0
Fugitive organic emissions from pump and compressor seals, valves, pipe flanges, pressure relief valves, and drains	1.5†	153	0	0	0	0	0	0	0	0
Total uncontrolled emissions excluding intermittent emissions	238,135	17,196	277,800	34,491	812	381	9,432	0	1,169-1,487	8,337-10,747
Total uncontrolled emissions including intermittent emissions	253,886	53,418	278,974	37,579	815	381	9,432	0	1,169-1,487	8,337-10,747

[*] N/A - not applicable

[†] **Estimated assuming an average molecular weight of 100.**

[‡] Streams with no data on the pollutant are not expected to contribute significantly to the total emission of the pollutant.

[**] Transient conditions would be expected twice per year for each of the four reactors, and up to 3 hours in duration each time.

[††] Estimated emission frequency is once a year, with an average emission duration of 24 hours.

[‡‡] Estimated emission frequency is once every four years, with an average emission duration of 24 hours.

[***] Transient conditions would be expected 30 times per year for each of the three units, and up to 3 hours in duration each time.

[†††] **These gas streams will normally be desulfurized and burned within the plant as fuel gas.**

As shown in Table 3-74, the Flexicoking gasifier/heater sour fuel gas (Stream 304) is the major contributor to hydrocarbon and CO emissions. This, of course, is an unlikely situation because Stream 304 will be used as fuel gas within the EDS plant. Other than the sour fuel gas, the two major contributors to hydrocarbon emissions are the slurry drier vent gas and the vacuum fractionator off-gas, both also contain sufficient heating value to be used as in-plant fuel gas. With these exceptions, hydrocarbon emissions are mostly contributed by fugitive emissions from pump and compressor seals, valves, pipe flanges, pressure relief valves and drains. Total uncontrolled hydrocarbon emissions were estimated to be 17,200 kg/hr if intermittent streams are excluded, and 53,400 kg/hr if all intermittent streams are included. If one excludes the gas streams which would likely be burned within the plant as fuel gas -- sour fuel gas (Stream 304), slurry drier vent gas (Stream 102), and vacuum fractionator off-gas (Stream 153) -- and if one excludes the intermittent streams, then uncontrolled hydrocarbon emissions would be only 218 kg/hr.

The acid gas from the DEA regenerator (Stream 508) is the largest source of sulfur emissions (in the form of H_2S) within the plant. Of course, in practice, this acid gas will be treated in a sulfur recovery system. If untreated, SO_2 emissions from the combustion of the sour fuel gas (Stream 304) would also be significant and greater than the uncontrolled SO_2 emissions from coal combustion in the steam and power plants. Total sulfur emissions, expressed as SO_2 equivalent,* were estimated to be 74,680 kg/hr if intermittent streams are excluded, and 80,500 kg/hr if intermittent streams are included. Of the 74,680 kg/hr of emissions of SO_2 equivalent, approximately 67% are contributed by the acid gas from the DEA regenerator.

Emissions of particulate matter were estimated to range from 8,340 to 10,750 kg/hr if intermittent streams are excluded. However, intermittent streams are not expected to contribute significantly to emissions of particulate matter. Particulate emissions are derived mainly from three sources: uncontrolled emissions from coal combustion in the steam plant (Stream 701a), uncontrolled emissions from coal combustion in the power plant (Stream 707a), and fugitive particulates from coal handling and crushing (Stream 013).

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* 1 kg of H_2S is equivalent to 1.8798 kg of SO_2; 1 kg of COS is equivalent to 1.0664 kg of SO_2.

3.4.1.2 MFS Case

The characteristics of uncontrolled gaseous waste streams for the EDS MFS case design are summarized in Table 3-75. The contributions of various streams to the total emission burden for the MFS case are similar to the contributions for the base case, excpet in cases where the waste streams are unique to the MFS case design. Contributions to the total emission burden from waste streams unique to the MFS design, however, are relatively insignificant except for the acid gas from the AGR unit in hydrogen purification (Stream 428).

For most pollutants listed in Table 3-75, emissions for the MFS case design are comparable to emissions for the base case design. The only exceptions are the considerably lower uncontrolled hydrocarbon and CO emissions for the MFS case design. This is because the sour fuel gas (Stream 304) for the MFS case design is only half the size of the sour fuel gas for the base case design.

3.4.2 Liquid Waste Streams

Mass discharge rates for liquid waste streams are summarized in this subsection according to process area. The total mass discharge rate and the mass discharge rates for nine pollutant parameters are presented. These nine pollutant parameters are: H_2S, NH_3, HCl, CO_2, phenols, organic acids, COD, TOC, and TDS. These are the most important parameters needed to evaluate pollution control options.

3.4.2.1 Base Case

The characteristics of uncontrolled aqueous waste streams for the EDS base case design are summarized in Table 3-76. There are eleven sour and phenolic wastewater streams in the base case design. The slurry drier cold separator wastewater (Stream 103) is a non-sour phenolic wastewater stream. The other ten sour and phenolic wastewater streams are Streams 106, 152, 155, 202, 252, 307, 308, 312, 430 and 431. Streams 430 and 431 would be expected to contain only trace levels of contaminants; Stream 430 only contains trace levels of CO_2 and Stream 431 contains 600 ppmv CO_2. In addition, Streams 451 and 452 contain NH_3 that require removal/recovery; although Stream 452 only

TABLE 3-75. SUMMARY OF UNCONTROLLED GASEOUS WASTE STREAMS FROM EDS COMMERCIAL PLANT (ILLINOIS COAL MFS CASE)

Uncontrolled Waste Stream	Total Flow, kmol/hr	Emission Rate, kg/hr								
		Total Hydrocarbon	CO	H_2S	NH_3	COS	SO_2	HCN	NO_x	Particulate Matter
Coal Preparation Process Area (Area 1)										
Stream 011 - fugitive dust from coal pile	N/A*	0	0	0	0	0	0	0	0	52
Stream 013 - fugitive particulates from coal handling and crushing	N/A	0	0	0	0	0	0	0	0	165-2,623
Coal Liquefaction Process Area (Area 2)										
Stream 102 - slurry drier vent gas	532	2,129	0	0.54	0	0	0	0	0	0
Stream 107 - flue gas from liquefaction slurry preheat furnace	34,535	2.4	14-20	0	0	0	91	0	97-186	4-12
Stream 803 - transient waste gas from liquefaction reactor (intermittent)**	3,280	30,836	426	2,849	2.7	0	0	0	0	0
Product Separation and Purification Process Area (Area 3)										
Streams 153/156 - vacuum fractionator off-gas†††	69	1,512	0	117	0	0	0	0	0	0
Stream 161 - flue gas from partial oxidation feed vacuum fractionator preheat furnaces	425	0.03	0.17-0.25	0	0	0	1.1	0	1.2-2.3	0.05-0.15
Stream 203 - flue gas from solvent hydrogenation fuel preheat furnaces	3,864	0.27	1.5-2.3	0	0	0	10	0	11-21	0.45-1.4
Stream 508 - acid gas from DEA regenerator	1,214	0	0	26,562	812	27	0	0	0	0
Liquefaction Residue Processing/Hydrogen Production Process Area (Area 4)										
Stream 304 - Flexicoking gasifier/heater sour fuel gas†††	27,225	6,677	132,643	4,124	0	150	0	0	0	0
Stream 428 - acid gas from acid gas removal unit in hydrogen purification	13,304	4.8	22	4,672	152	30	0	0	0	0
Stream 438 - hydrogen plant deaerator vent	6.2	0	<0.5	0	0	0	0	0	0	0
Stream 440 - flash gas from partial oxidation unit	254	1.3	2,176	382	0.68	18	0	0	0	0
Stream 449/450 - regeneration/decommissioning off-gas from high and low temperature shift catalysts (intermittent)††	22,255	0	0	0	0	0	15,683	0	0	608

(Continued)

TABLE 3-75. (Continued)

Uncontrolled Waste Stream	Total Flow, kmol/hr	Emission Rate, kg/hr								
		Total Hydrocarbon	CO	H_2S	NH_3	COS	SO_2	HCN	NO_x	Particulate Matter
(Area 4) (Continued)										
Stream 801 - transient waste gas from Flexicoking (intermittent)††	511	4,372	566	262	126	0	0	0	0	0
Stream 802 - transient waste gas from partial oxidation units (intermittent)***	12,121	390	60,242	1,244	36	186	0	2.6	0	0
Auxiliary Operations Process Area (Area 5)										
Stream 701a - flue gas from steam generation system	69,150	16	59 - 65	0	0	0	6,528	0	670 - 750	5,741
Stream 707a - flue gas from power generation system	26,110	10	34	0	0	0	4,775	0	431	4,245
Stream 731 - drift and evaporation from cooling tower	N/A	No data‡	0	0	0	0	0	0	0	No data
Stream 751 - evaporative emissions from product/byproduct storage	0.02†	2.2	0	0	0	0	0	0	0	0
Fugitive organic emissions from pump and compressor seals, valves, pipe flanges, pressure relief valves, and drains	1.5†	153	0	0	0	0	0	0	0	0
Total uncontrolled emissions excluding intermittent emissions	176,690	10,508	134,957	35,858	965	225	11,405	0	896 - 1,076	10,208-12,675
Total uncontrolled emissions including intermittent emissions	214,857	46,106	196,190	40,213	1,129	411	27,088	2.6	1,210-1,390	10,816-13,283

*N/A - not applicable.

†Estimated assuming an average molecular weight of 100.

‡Not expected to contribute significantly to the total emissions of the pollutant.

**Transient conditions would be expected twice per year for each of the four reactors, and up to 3 hours in duration each time.

††Estimated emission frequency of once a year, with emission duration from 12 to 72 hours.

‡‡Transient conditions would be expected 30 times per year for each of the two units, and up to 3 hours in duration each time.

***Transient conditions expected once a year, and only to one of the partial oxidation units and up to 24 hours in duration.

†††**These gas streams will normally be desulfurized and burned within the plant as fuel gas.**

TABLE 3-76. SUMMARY OF UNCONTROLLED LIQUID WASTE STREAMS FROM EDS COMMERCIAL PLANT (ILLINOIS COAL BASE CASE)

Uncontrolled Waste Stream	Total Discharge Rate, kg/hr	Discharge Rate, kg/hr								
		H_2S	NH_3	HCl	CO_2	Phenols	Organic Acids	COD	TOC	TDS
Coal Preparation Process Area (Area 1)										
Stream 012 - coal pile runoff (intermittent)	17,333	0	0	0	0	0	0	Low[*]	Low[*]	139[†]
Coal Liquefaction Process Area (Area 2)										
Stream 103 - slurry drier cold separator wastewater	155,640	0.6	12	2	46	151	0	No data	No data	No data[‡]
Stream 106 - liquefaction cold separator wastewater	100,583	3,251	2,297	249	1,956	1,380	745	5,933	1,950	810
Product Separation and Purification Process Area (Area 3)										
Stream 152 - atmospheric fractionator overhaul drum wastewater	29,105	122	157	37	9.5	536	91	3,280	1,024	5.4
Stream 155 - vacuum fractionator wastewater	4,839	41	0.2	0.1	3.5	24	1.6	21	6.3	0.7
Stream 202 - solvent hydrogenation cold separator wastewater	12,561	730	684	0	0	13	0	No data	No data	No data
Stream 252 - solvent hydrogenation fractionator overhead drum wastewater	12,461	333	232	0	0	64	0	678	131	0.4
Liquefaction Residue Processing/Hydrogen Production Process Area (Area 4)										
Stream 307 - Flexicoking recontacting drum wastewater	1,312	23	31	0	34	9.5	0	No data	No data	No data
Stream 308 - Flexicoking fractionator overhead drum wastewater	99,910	177	214	15	261	389	0	No data	No data	No data
Stream 312 - Flexicoking heater overhead drum wastewater	28,400	171	396	0	616	0.5	0	No data	No data	No data
Stream 403 - knockout drum wastewater in H_2 cryogenic recovery	1,334	0	0	0	0	0	0	0	0	No data
Stream 430 - blowdown and K.O. drum wastewater from hydrogen generation	55,430	0	0	0	0	0	0	0	0	No data

(Continued)

TABLE 3-76. (Continued)

Uncontrolled Waste Stream	Total Discharge Rate, kg/hr	Discharge Rate, kg/hr								
		H_2S	NH_3	HCl	CO_2	Phenols	Organic Acids	COD	TOC	TDS
Area 4 (Continued)										
Stream 431 - Catacarb overhead receiver wastewater in hydrogen generation	22,600	0	0	0	**14**	0	0	0	0	No data
Stream 451 - aqueous ammonia from ammonia synthesis	22,560	0	2,600	0	0	0	0	0	0	No data
Stream 452 - knockout drum wastewater in ammonia synthesis	454	0	0	0	0	0	0	0	0	No data
Auxiliary Operations Process Area (Area 5)										
Stream 702 - blowdown from steam generation system	67,000	0	0	0	0	0	0	Low	Low	268
Stream 708 - blowdown from power generation system	2,800	0	0	0	0	0	0	Low	Low	0.6 - 1.4
Stream 723 - regeneration wastes from water demineralization	30,000	0	0	0	0	0	0	Low	Low	494**
Stream 732 - cooling tower blowdown	630,000	0	0	0	0	0	0	Low	Low	315
Low quality rain runoff (intermittent)	318,000	0	0	0	0	No data	No data	No data	No data	No data
Total uncontrolled discharge excluding intermittent discharges	1,277,079	4,849	6,623	303	2,940	2,567	838	**No data**	**No data**	1,755
Total uncontrolled discharge including intermittent discharges	1,612,412	4,849	6,623	303	2,940	2,567	838	**No data**	**No data**	1,894

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[*] Typical COD and TOC values for Streams 012, 702, 708, 723, and 732 are below 100 mg/l (36).

[†] TDS for coal pile runoff was computed assuming a TDS value of 8,000 mg/l. TDS values for coal pile runoff can vary over a very wide range (e.g., 3,000 to 20,000 mg/l) (36).

[‡] Although no data are publicly available, TDS for Streams 103, 202, 307, 308, 312, 403, 430, 431, 452, are expected to be low.

[**] The TDS value for this waste stream was estimated from the concentrations of sulfate, nitrate, and chloride compounds present to be 16,460 mg/l.

contains trace levels of NH_3. These thirteen wastewater streams from Areas 2, 3 and 4 are unique to the EDS process and contribute a major portion of all pollutant loadings except TDS. Discharges from auxiliary operations and intermittent runoff streams, although comparably larger in magnitude, are expected to contribute mainly to the TDS loading of wastewater streams from the EDS process.

3.4.2.2 MFS Case

The characteristics of uncontrolled aqueous waste streams for the EDS MFS case design are summarized in Table 3-77. There are also eleven sour and phenolic wastewater streams in the MFS case design. As in the base case, Stream 103 is a non-sour phenolic wastewater stream. The other ten sour and phenolic wastewater streams are Streams 106, 152, 155/157, 202, 252, 307, 308, 312, 430 and 441. Of these streams, Stream 430 would be expected to be relatively free of contaminants and contain only trace levels of H_2S, NH_3, and CO_2. The slag filtrate from the partial oxidation unit (Stream 443) is expected to contain low levels of H_2S and phenols. Again, wastewater streams from Areas 2, 3 and 4 are unique to the EDS design and would have major impact on the selection of wastewater treatment processes. Wastewater streams from auxiliary operations and intermittent runoff streams only have impact on the total flow of wastewater and the TDS loading.

3.4.3 Solid Waste Streams

Mass discharge rates for solid wastes are summarized in this subsection in tabular form. The presentation of the solid waste summary, however, is different from the summary for gaseous and liquid waste streams in two aspects. First, composition data and leaching characteristics for solid wastes are presented in a more descriptive format. This is because quantitative composition data for most solid wastes are not available, and leaching characteristics are only available for three solid wastes. Second, the generation rate for solid wastes is expressed as Mg/yr; no hourly rate is given. This is because only six solid wastes are generated on a continuous basis and an annual rate provides a better indication of the solid waste disposal/recovery requirements.

TABLE 3-77. SUMMARY OF UNCONTROLLED LIQUID WASTE STREAMS FROM EDS COMMERCIAL PLANT (ILLINOIS COAL MFS CASE)

Uncontrolled Waste Stream	Total Discharge Rate, kg/hr	Discharge Rate, kg/hr								
		H_2S	NH_3	HCl	CO_2	Phenols	Organic Acids	COD	TOC	TDS
Coal Preparation Process Area (Area 1)										
Stream 012 - coal pile runoff (intermittent)	17,708	0	0	0	0	0	0	Low*	Low*	142†
Coal Liquefaction Process Area (Area 2)										
Stream 103 - slurry drier cold separator wastewater	155,640	0.6	12	2	46	151	0	**No data**	**No data**	No data‡
Stream 106 - liquefaction cold separator wastewater	100,583	3,251	2,297	249	1,956	1,380	745	**5,933**	**1,950**	810
Product Separation and Purification Process Area (Area 3)										
Stream 152 - atmospheric fractionator overhead drum wastewater	29,105	122	157	37	9.5	536	91	**3,280**	**1,024**	5.4
Stream 155/157 - vacuum fractionator wastewater	4,839	41	0.2	0.1	3.5	24	1.6	**21**	**6.3**	0.7
Stream 202 - solvent hydrogenation cold separator wastewater	12,561	730	684	0	0	13	0	**No data**	**No data**	No data
Stream 252 - solvent hydrogenation fractionator overhead drum wastewater	12,461	333	232	0	0	64	0	**678**	**131**	0.4
Liquefaction Residue Processing/Hydrogen Production Process Area (Area 4)										
Stream 307 - Flexicoking recontacting drum wastewater	656	12	16	0	17	4.8	0	**No data**	**No data**	No data
Stream 308 - Flexicoking fractionator overhead drum wastewater	49,955	89	107	7.3	131	194	0	**No data**		No data
Stream 312 - Flexicoking heater overhead drum wastewater	14,200	86	198	0	308	0.2	0	**No data**	**No data**	No data
Stream 403 - knockout drum wastewater in H_2 cryogenic recovery	1,334	0	0	0	0	0	0	0	0	No data
Stream 430 - blowdown and knockout drum wastewater from hydrogen generation	3,110	0	0	0	0	0	0	0	0	No data

(Continued)

TABLE 3-77. (Continued)

Uncontrolled Waste Stream	Total Discharge Rate, kg/hr	Discharge Rate, kg/hr								
		H_2S	NH_3	HCl	CO_2	Phenols	Organic Acids	COD	TOC	TDS
Area 4 (Continued)										
Stream 441 - sour water from the partial oxidation unit	45,550	3-15	56-123	No data	No data	0.001	No data	18-35	10-35	19-91
Stream 443 - slag filtrate from the partial oxidation unit	9,320	<0.06	1.2-3.5	No data	No data	<0.0001	0.45-2.1	0.26-0.35	0.26-0.35	0.26-3.3
Auxiliary Operations Process Area (Area 5)										
Stream 702 - blowdown from the steam generation system	55,000	0	0	0	0	0	0	Low	Low	220
Stream 708 - blowdown from the power generation system	2,000	0	0	0	0	0	0	Low	Low	1
Stream 723 - regeneration wastes from water demineralization	34,000	0	0	0	0	0	0	Low	Low	560
Stream 732 - cooling tower blowdown	510,000	0	0	0	0	0	0	Low	Low	255
Low quality rain runoff (intermittent)	318,000	0	0	0	0	No data	No data	No data	No data	No data
Total uncontrolled discharge excluding intermittent discharges	1,040,314	4,674	3,795	295	2,471	2,367	838	No data	No data	1,909
Total uncontrolled discharge including intermittent discharges	1,376,022	4,674	3,795	295	2,471	2,367	838	No data	No data	2,051

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[*] Typical COD and TOC values for streams 012, 702, 708, 723 and 732 are below 100 mg/l (36).

[+] TDS for coal pile runoff was computed assuming a TDS value of 8,000 mg/l. TDS values for coal pile runoff can vary over a very wide range (e.g., 3,000 - 20,000 mg/l) (36).

[†] Although no data are publicly available, TDS for Streams 103, 202, 307, 308, 312, 403, and 430 are expected to be low.

[**] The TDS value for this waste stream was estimated from the concentrations of sulfate, nitrate, and chloride compounds present to be 16,460 mg/l.

3.4.3.1 Base Case

The characteristics of uncontrolled solid waste streams for the EDS base case design are presented in Table 3-78. As shown in this table, the four solid wastes from Flexicoking (Streams 302, 303, 306 and 313) are responsible for 97.2 wt % of the total uncontrolled solid wastes generated for the base case design. There are four other major solid waste streams if air pollution controls are installed on the steam and power plants, as would be expected. These four solid waste streams are: fly ash from steam generation system, FGD sludge from steam generation system, fly ash from power generation system, and FGD sludge from power generation system. Sludge generated in wastewater treatment is a major solid waste stream from water pollution control.

3.4.3.2 MFS Case

The characteristics of uncontrolled solid waste streams for the EDS base case design are presented in Table 3-79. For the MFS case, the four solid wastes from Flexicoking (Streams 302, 303, 306 and 313) together with the slag from the partial oxidation unit (Stream 442) are responsible for 96.9 wt % of the total uncontrolled solid wastes generated. As in the base case design, there are four other major solid wastes if flue gas emissions from steam and power plants are controlled, and a large sludge waste stream from wastewater treatment.

3.4.4 Product Streams

The product rates and composition of EDS products are summarized in this subsection. Since only limited product composition data that address individual compounds or compound classes are available, this information is summarized in a descriptive format. By-products recovered from pollution control, such as crude phenol, sulfur, and ammonia are not covered here.

3.4.4.1 Base Case

The characteristics of product streams for the EDS base case design are presented in Table 3-80. Total production rate of C_3^+ liquids is 10,493 m^3 (66,000 bbl or 60,240 BFOE*) per stream day. Product composition data on naphtha and blended fuel oil are based on very limited and preliminary sampling and analysis efforts.

TABLE 3-78. SUMMARY OF UNCONTROLLED SOLID WASTES FROM EDS COMMERCIAL PLANT (ILLINOIS COAL BASE CASE)

Uncontrolled Waste Stream	Generation Rate, Mg/yr	Composition Data/Leaching Characteristics
Coal Liquefaction Process Area (Area 2)		
Stream 108 - solids accumulated in slurry drier	820	Similar in composition to feed coal.
Product Separation and Purification Process Area (Area 3)		
Stream 204 - spent solvent hydrogenation catalyst (Ni-Mo)	821	Contains nickel, molybdenum, carbonaceous material, and sulfides.
Liquefaction Residue Processing/Hydrogen Production Process Area (Area 4)		
Stream 302 - Flexicoking gasifier/heater dry fines	137,950	Contains 80 wt % coal ash and 20 wt % carbonaceous material. Concentrations of leachate using EPA EP in ppmw: As <0.055; Ba - 1.76; Cd - 0.0425; Cr - 0.0042; Pb - 0.0462; Hg - 0.0004; Se <0.036; Ag <0.002.
Stream 303 - Flexicoking gasifier/heater wet fines	344,870*	Also contains 80 wt % coal ash and 20 wt % carbonaceous material for the solids portion.
Stream 306 - Flexicoking gasifier/heater bed coke	707,100	Contains 61 wt % coal ash and 39 wt % carbonaceous material. Concentrations of leachate using EPA EP in ppmw: As <0.036; Ba - 0.032; Cd - <0.0066; Cr - <0.003; Pb - <0.019; Hg - 0.0009; Se - <0.055; Ag - <0.0026.
Stream 313 - Flexicoking heater/reactor chunks/agglomerates	19,860	Concentrations of leachinate using EPA EP in ppmw: As <0.055; Ba - 0.0407; Cd <0.0048; Cr <0.0006; Pb <0.023; Hg <0.0002; Se <0.036; Ag <0.002.
Stream 404 - spent hydrotreater catalyst from cryogenic hydrogen recovery (Ni-Mo)	68.5	Contains nickel, molybdenum, carbonaceous material, and sulfides.
Stream 405 - spent drying agents from cryogenic hydrogen recovery (zeolite)	263	Similar in composition to zeolites (hydrated aluminosilicates)
Stream 433 - spent sulfur guard from hydrogen generation (ZnO)	271	Contains zinc oxide, zinc sulfide, and carbonaceous material.
Stream 435 - spent reformer catalyst from hydrogen generation (Ni-U)	181	Contains 13 wt % Ni, 12.1 wt % U, 0.3 wt % K, alumina, carbonaceous material and trace of sulfides.
Stream 436 - spent shift catalyst from hydrogen generation (iron oxide)	321	Contains iron oxide and carbonaceous material.
Stream 439 - spent methanation catalyst from hydrogen generation (NiO)	115	Contains nickel oxide and carbonaceous material.
Stream 453 - spent drying agents from ammonia synthesis (alumina)	172	Contains alumina and traces of ammonia.
Stream 454 - spent ammonia synthesis catalyst (iron/iron oxide)	166	Contains mainly iron and iron oxides, 5-8% total of Al_2O_3, K_2O, CaO and MgO, traces of NH_3 and ammonium compounds.
Auxiliary Operations Process Area (Area 5)		
Stream 704 - bottom ash from steam generation system	3,900	Similar in composition to coal ash.
Stream 710 - bottom ash from power generation system	10,740	Similar in composition to coal ash.
Stream 722 - raw water treatment sludge	15,120*	Contains 36.4 wt % $CaCO_3$, 3.6 wt % $Mg(OH)_2$, and 60 wt % water.
Total uncontrolled solid waste	1,242,799	

*40 wt % solids basis.

TABLE 3-79. SUMMARY OF UNCONTROLLED SOLID WASTES FROM EDS COMMERCIAL PLANT (ILLINOIS COAL MFS CASE)

Uncontrolled Waste Stream	Generation Rate, Mg/yr	Composition Data/Leaching Characteristics
Coal Liquefaction Process Area (Area 2)		
Stream 108 - solids accumulated in slurry drier	820	Similar in composition to feed coal.
Product Separation and Purification Process Area (Area 3)		
Stream 204 - spent solvent hydrogeneration catalyst	821	Contains nickel, molybdenum, carbonaceous material, and sulfides.
Liquefaction Residue Processing/Hydrogen Production Process Area (Area 4)		
Stream 302 - Flexicoking gasifier/heater dry fines	65,670	Contains 80 wt % coal ash and 20 wt % carbonaceous material. Concentrations of leachate using EPA EP in ppmw: As <0.055; Ba - 1.76; Cd - 0.0425; Cr - 0.0042; Pb - 0.0462; Hg - 0.0004; Se <0.036; Ag <0.002.
Stream 303 - Flexicoking gasifier/heater wet fines	164,160*	Also contains 80 wt % coal ash and 20 wt % carbonaceous materials for the solids portion.
Stream 306 - Flexicoking gasifier/heater bed coke	363,030	Contains 61 wt % coal ash and 39 wt % carbonaceous material. Concentrations of leachate using EPA EP in ppmw: As <0.036; Ba - 0.032; Cd - <0.0066; Cr <0.003; Pb <0.019; Hg - 0.0009; Se <0.055; Ag <0.0026.
Stream 313 - Flexicoking heater/reactor chunks/agglomerates	9,660	Concentrations of leachate using EPA EP in ppmw: As <0.055; Ba - 0.0407; Cd <0.0048; Cr <0.0006; Pb <0.023; Hg <0.0002; Se <0.036; Ag <0.002.
Stream 404 - spent hydrotreater catalyst from cryogenic hydrogen recovery (Ni-Mo)	37.2	Contains nickel, molybdenum, carbonaceous material, and sulfides.
Stream 405 - spent drying agents from cryogenic hydrogen recovery (zeolite)	177	Similar in composition to zeolites (hydrated aluminosilicates).
Stream 442 - slag from partial oxidation unit	447,300	Similar in composition to coal ash with up to 2 wt % carbonaceous material.
Stream 444 - spent high temperature shift catalyst from hydrogen generation (Co-Mo)	451	Contains cobalt, molybdenum, carbonaceous material, sulfides, sulfur, and volatile trace metals.
Stream 445 - spent low temperature shift catalyst from hydrogen generation (Co-Mo)	254	Contains cobalt, molybdenum, carbonaceous material, sulfides, and possibly volatile trace metals.
Auxiliary Operations Process Area (Area 5)		
Stream 704 - bottom ash from steam generation system	10,400	Similar in composition to coal ash.
Stream 710 - bottom ash from power generation system	7,700	Similar in composition to coal ash.
Stream 722 - raw water treatment sludge	13,477†	Contains 36.4 wt % CaCO, 3.6 wt % $Mg(OH)_2$, and 60 wt % water.
Total uncontrolled solid waste	1,083,957	

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*40 wt % solids basis.

†79 wt % solids basis.

3.4.4.2 MFS Case

The characteristics of product streams for the EDS MFS case design are presented in Table 3-81. For this design, total production of C_3^+ liquids amounts to 10,392 m^3 (65,367 bbl or 59,180 BFOE) per stream day. However, total production of hydrocarbon products when pipeline gas is included (as barrels fuel oil equivalent) increases to 11,300 m^3 (71,080 bbl) of fuel oil equivalent per stream day, or 18% higher than the base case design. Composition of products for the MFS case design are either similar or identical to the composition of the same products for the base case design, except pipeline gas is not produced in the base case.

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* 1 barrel fuel oil equivalent (BFOE) = 6,050,000 Btu (LHV)

TABLE 3-80. SUMMARY OF PRODUCTS FROM EDS COMMERCIAL PLANT (ILLINOIS COAL BASE CASE)

Product	Production Rate			Product Composition
	m^3/SD	bbl/SD	BFOE*/SD	
C_3 LPG	464	2,915	1,780	Contains 95.8 wt % C_3, 0.9 wt % C_2^-, and 3.3 wt % C_4^+
C_4 LPG	392	2,465	1,680	Contains 95.1 wt % C_4, 3.2 wt % C_3^-, and 1.7 wt % C_5^+
Naphtha	3,770	23,710	20,100	Contains 85.20 wt % C, 13.16 wt % H, 0.43 wt % S, 0.06 wt % N, and 1.15 wt % O. Analysis of a naphtha sample from the solvent fractionation tower indicated: 80 wt % of total organics are aliphatic hydrocarbons, 11 wt % are benzenes and substituted benzenes, and 0.7 wt % are phenols.
Blended fuel oil	5,868	36,910	36,680	Contains 87.56 wt % C, 8.83 wt % H, 0.54 wt % S, 0.77 wt % N, and 2.30 wt % O. Preliminary analysis of a light solvent fuel oil indicated that the major organic compounds are aliphatic hydrocarbons, benzenes and substituted benzenes, phenols, and heterocyclic oxygen compounds such as furans and especially C_1-C_5 benzofurans.
Total	10,493	66,000	60,240	

* 1 BFOE (barrel fuel oil equivalent) = 6,050,000 Btu (LHV).

TABLE 3-81. SUMMARY OF PRODUCTS FROM EDS COMMERCIAL PLANT (ILLINOIS COAL MFS CASE)

Product	Production Rate			Product Composition
	m^3/SD	bbl/SD	BFOE*/SD	
Pipeline gas	1,991,000 Nm^3/SD	70.3 MMSCFD	11,900	Contains 63.0 vol. % CH_4, 22.3 vol. % C_2, 2.7 vol. % C_3^+, 5.4 vol. % N_2, 3.4 vol. % H_2, and 3.2 vol. % CO.
C_3 LPG	667	4,198	2,570	Contains 95.8 wt % C_3, 0.9 wt % C_2^-, and 3.3 wt % C_4^+.
C_4 LPG	405	2,548	1,740	Contains 95.1 wt % C_4, 3.2 wt % C_3^-, and 1.7 wt % C_5^+.
Naphtha	3,695	23,243	19,710	Contains 85.20 wt % C, 13.16 wt % H, 0.43 wt % S, 0.06 wt % N, and 1.15 wt % O. Analysis of a naphtha sample from the solvent fractionation tower indicated: 80 wt % of total organics are aliphatic hydrocarbons, 11 wt % are benzenes and substituted benzenes, and 0.7 wt % are phenols.
Blended fuel oil	5,625	35,378	35,160	Contains 87.72 wt % C, 8.89 wt % H, 0.51 wt % S, 0.75 wt % N, and 2.13 wt % O. Preliminary analysis of a light solvent fuel oil indicated that the major organic components are aliphatic hydrocarbons, benzenes and substituted benzenes, phenols, and heterocyclic oxygen compounds such as furans and especially C_1-C_5 benzofurans.
Total excluding gas	10,392	65,367	59,180	
Total including pipeline gas	--	--	71,080	

* 1 BFOE = 6,050,000 Btu (LHV).

3.5 WASTE/CONTROL TECHNOLOGY INDEX

The preceding parts of this section have provided a general description of base case and MFS case designs of the EDS process, and test data and engineering estimates characterizing the uncontrolled or primary waste streams expected. Section 4 of this manual presents information on the available control techniques for these primary waste streams and illustrative examples of both individual control technologies and integrated systems of control technologies applied to specific streams. As will be discussed in Section 4, residuals or secondary waste streams are generated as a result of the application of some control technologies; control of secondary waste streams is discussed in conjunction with the illustrative examples in Section 4.

To aid users in locating characterization data and control technology information for any waste stream addressed in this manual, a cross reference index was developed. This index is presented in Table 3-82 and indicates where characterization data can be found in Section 3 and where control technology information can be found in Section 4 for each primary waste stream. The waste streams in Table 3-82 are grouped by the process area from which they originate and then further grouped within each operation by waste medium. Similar types of information on secondary waste streams are presented in Table 3-83. The entries in Table 3-83 are not meant to imply that those streams will necessarily be found in EDS commercial plants, but that if the control techniques listed are used, then those streams will be produced.

TABLE 3-82. CROSS-REFERENCE INDEX FOR PRIMARY WASTE STREAMS

Process Area/Media/Waste Stream	PCTM Section Reference for Waste Characterization	PCTM Section Reference for Control Technique
Coal Preparation Process Area		
Gaseous Waste Streams		
Fugitive dust from coal pile (Stream 011)	3.3.1.1.1	4.2.4
Fugitive particulate from coal handling and crushing (Stream 013)	3.3.1.2.1	4.2.6
Liquid Waste Stream		
Coal pile runoff (Stream 012)	3.3.1.1.2	4.3.2
Coal Liquefaction Process Area		
Gaseous Waste Streams		
Slurry drier vent gas (Stream 102)	3.3.2.1.1	4.2.1.4
Flue gas from liquefaction slurry preheat furnace (Stream 107)	3.3.2.1.1	4.2.2.2
Transient gas from liquefaction reactor (Stream 803)	3.3.2.1.1	4.2.1.5
Liquid Waste Streams		
Slurry drier cold separator wastewater (Stream 103)	3.3.2.1.2	4.3.1.4, 4.3.1.6
Liquefaction cold separator wastewater (Stream 106)	3.3.2.2.2	4.3.1.1, 4.3.1.6
Solid Waste Stream		
Solids accumulated in the slurry drier (Stream 108)	3.3.2.1.3	4.4.2.3
Product Separation and Purification Process Area		
Gaseous Waste Streams		
Vacuum fractionator off-gas (Stream 153)	3.3.3.1.1	4.2.1.4
Partial oxidation feed vacuum fractionator off-gas (Stream 156)	3.3.3.1.1	4.2.1.4
Flue gas from partial oxidation feed vacuum fractionator preheat furnaces (Stream 161)	3.3.3.1.1	4.2.2.2

(continued)

TABLE 3-82. (Continued)

Process Area/Media/Waste Stream	PCTM Section Reference for Waste Characterization	PCTM Section Reference for Control Technique
Gaseous Waste Streams (continued)		
Flue gas from solvent hydrogenation fuel product furnaces (Stream 203)	3.3.3.2.1	4.2.2.2
Acid gas from DEA unit (Stream 508)	3.3.3.3.1	4.2.1.1, 4.2.1.6
Liquid Waste Streams		
Atmospheric fractionator overhead drum wastewater (Stream 152)	3.3.3.1.2	4.3.1.1, 4.3.1.6
Vacuum fractionator wastewater (Stream 155)	3.3.3.1.2	4.3.1.1, 4.3.1.6
Partial oxidation feed vacuum fractionator wastewater (Stream 157)	3.3.3.1.2	4.3.1.1, 4.3.1.6
Solvent hydrogenation coal separator wastewater (Stream 202)	3.3.3.2.2	4.3.1.1, 4.3.1.6
Solvent hydrogenation fractionator overhead drum wastewater (Stream 252)	3.3.3.2.2	4.3.1.1, 4.3.1.6
Solid Waste Stream		
Spent solvent hydrogenation catalyst (Stream 204)	3.3.3.2.3	4.4.5.1, 4.4.6.1
Liquefaction Residue Processing/Hydrogen Production Process Area		
Gaseous Waste Streams		
Flexicoking gasifier/heater sour gas (Stream 304)	3.3.4.1.1	4.2.1.3
Vent gas from CO_2 removal (Stream 426)	3.3.4.2.1	4.2.3.1
Acid gas from acid gas removal unit in hydrogen purification (Stream 428)	3.3.4.2.1	4.2.1.2
Flue gas from hydrogen plant reformer furnaces (Stream 434)	3.3.4.2.1	4.2.2.2
Hydrogen plant deaerator vent (Stream 438)	3.3.4.2.1	4.2.3.2
Flash gas from partial oxidation unit (Stream 440)	3.3.4.2.1	4.2.1.1, 4.2.1.6
Regeneration/decommissioning off-gas from reformer catalyst (Stream 446)	3.3.4.2.1	4.2.3.3
Decommissioning off-gas from methanation catalyst (Stream 448)	3.3.4.2.1	4.2.3.3

(continued)

TABLE 3-82. (Continued)

Process Area/Media/Waste Stream	PCTM Section Reference for	
	Waste Characterization	Control Technique
Gaseous Waste Streams (continued)		
Regeneration/decommissioning off-gas from high temperature shift conversion catalyst (Stream 449)	3.3.4.2.1	4.2.2.3
Regeneration/decommissioning off-gas from low temperature shift conversion catalyst (Stream 450)	3.3.4.2.1	4.2.2.3
Transient gas from Flexicoking (Stream 801)	3.3.4.1.1	4.2.1.5
Transient waste gas from partial oxidation unit (Stream 802)	3.3.4.2.1	4.2.1.5
Liquid Waste Streams		
Flexicoking recontacting drum wastewater (Stream 307)	3.3.4.1.2	4.3.1.1, 4.3.1.6
Flexicoking fractionator overhead drum wastewater (Stream 308)	3.3.4.1.2	4.3.1.1, 4.3.1.6
Flexicoking heater overhead drum wastewater (Stream 312)	3.3.4.1.2	4.3.1.2, 4.3.1.6
Knockout drum wastewater in H_2 cryo recovery (Stream 403)	3.3.4.2.2	4.3.1.2, 4.3.1.6
Blowdown and K.O. drum wastewater from hydrogen generation (Stream 430)	3.3.4.2.2	4.3.1.2, 4.3.1.6
Catacarb overhead received wastewater in hydrogen generation (Stream 431)	3.3.4.2.2	4.3.1.2, 4.3.1.6
Sour water from partial oxidation unit (Stream 441)	3.3.4.2.2	4.3.1.2, 4.3.1.6
Slag filtrate from partial oxidation unit (Stream 443)	3.3.4.2.2	4.3.1.5, 4.3.1.6
Aqueous ammonia from ammonia synthesis (Stream 451)	3.3.4.2.2	4.3.1.3, 4.3.1.6
Knockout drum wastewater in ammonia synthesis (Stream 452)	3.3.4.2.2	4.3.1.2, 4.3.1.6
Solid Waste Streams		
Flexicoking gasifier/heater dry fines (Stream 302)	3.3.4.1.3	4.4.2.1, 4.4.6.1, 4.4.6.2
Flexicoking gasifier/heater wet fines (Stream 303)	3.3.4.1.3	4.4.2.1, 4.4.6.1, 4.4.6.2
Flexicoking gasifier/heater bed coke (Stream 306)	3.3.4.1.3	4.4.2.1, 4.4.6.1, 4.4.6.2
Flexicoking heater/reactor chunks/agglomerates (Stream 313)	3.3.4.1.3	4.4.2.2, 4.4.6.1

(continued)

TABLE 3-82. (Continued)

Process Area/Media/Waste Stream	PCTM Section Reference for Waste Characterization	PCTM Section Reference for Control Technique
Solid Waste Streams		
Spent hydrotreater catalyst in H_2 cryo recovery (Stream 404)	3.3.4.2.3	4.4.5.1, 4.4.6.1
Spent drying agents in H_2 cryo recovery (Stream 405)	3.3.4.2.3	4.4.5.1, 4.4.6.1
Spent sulfur guard in hydrogen generation (Stream 433)	3.3.4.2.3	4.4.5.1, 4.4.6.1
Spent reformer catalyst in hydrogen generation (Stream 435)	3.3.4.2.3	4.4.5.1, 4.4.6.1
Spent shift catalyst in hydrogen generation (Stream 436)	3.3.4.2.3	4.4.5.1, 4.4.6.1
Spent methanation catalyst in hydrogen generation (Stream 439)	3.3.4.2.3	4.4.5.1, 4.4.6.1
Slag from partial oxidation unit (Stream 442)	3.3.4.2.3	4.4.2.4, 4.4.6.1
Spent high temperature shift catalyst in hydrogen generation (Stream 444)	3.3.4.2.3	4.4.5.1, 4.4.6.1
Spent low temperature shift catalyst in hydrogen generation (Stream 445)	3.3.4.2.3	4.4.5.1, 4.4.6.1
Spent drying agents in ammonia synthesis (Stream 453)	3.3.4.2.3	4.4.5.1, 4.4.6.1
Spent ammonia synthesis catalyst (Stream 454)	3.3.4.2.3	4.4.5.1, 4.4.6.1
Auxiliary Operations Process Area		
Gaseous Waste Streams		
Flue gas from steam generation system (Stream 701a)	3.3.5.2.1	4.2.2.1
Flue gas from power generation system (Stream 707a)	3.3.5.2.1	4.2.2.1
Drift and evaporation from cooling tower (Stream 731)	3.3.5.3.1	not evaluated*
Evaporative emissions from product and byproduct storage (Stream 751)	3.3.5.5.1	4.2.5.1
Liquid Waste Streams		
Blowdown from steam generation system (Stream 702)	3.3.5.2.2	4.3.2
Blowdown from power generation system (Stream 708)	3.3.5.2.2	4.3.2
Regeneration wastes from water demineralization (Stream 723)	3.3.5.1.2	4.3.2
Cooling tower blowdown (Stream 732)	3.3.5.3.2	4.3.2

(continued)

TABLE 3-82. (Continued)

Process Area/Media/Waste Stream	PCTM Section Reference for	
	Waste Characterization	Control Technique
Solid Waste Streams		
Bottom ash from steam generation system (Stream 704)	3.3.5.2.3	4.4.2.5, 4.4.6.1
Bottom ash from power generation system (Stream 710)	3.3.5.2.3	4.4.2.5, 4.4.6.1
Raw water treatment sludge (Stream 722)	3.3.5.1.3	4.4.2.8, 4.4.6.1
Fugitive Emissions		
Gaseous Waste Stream		
Fugitive organic emissions from process equipment	3.3.7.1	4.2.5.2
Liquid Waste Stream		
Low quality rain runoff	3.3.7.2	not evaluated+

*not evaluated - drift loss is part of the cooling tower design, and evaporative loss is a function of the cooling tower operating Δ T.

+not evaluated - but treatment for this wastewater is similar to treatment of Source Type 2 wastewater streams.

TABLE 3-83. CROSS-REFERENCE INDEX FOR SECONDARY WASTE STREAMS FROM POLLUTION CONTROL OPERATIONS

	PCTM Section Reference for		
Control Technique/Secondary Waste	Waste Characterization	Control Technique	Appendix
Air Pollution Control			
Claus Process			
Spent catalyst	4.2.1.1.1.1	4.4.5.1	A-6
Stretford Process			
Oxidizer vent gas	4.2.1.2.1.1	4.2.1.2.1.1	A-7
Solution purge	4.2.1.2.1.1	4.2.1.2.1.1	A-7
Beavon Sulfur Removal Process (BSRP)			
Oxidizer vent gas	4.2.1.1.2.1	4.2.1.2.1.1	A-9
Sour condensate	4.2.1.1.2.1	not evaluated*	A-9
Stretford solution purge	4.2.1.1.2.1	4.2.1.2.1.1	A-9
Spent catalyst	4.2.1.1.2.1	4.4.5.1	A-9
Shell Claus Off-Gas Treating (SCOT) Process			
Sour condensate	4.2.1.1.2.2	not evaluated	A-8
Spent catalyst	4.2.1.1.2.2	4.4.5.1	A-8
Wellman-Lord Process			
Acidic wastewater	4.2.1.1.2.3	not evaluated	A-10
Thiosulfate/sulfate purge	4.2.1.1.2.3	not evaluated	A-10
Electrostatic Precipitators			
Boiler fly ash	4.2.2.1.2.1	4.4.2.6	A-13
Limestone Scrubbing			
FGD sludge	4.2.2.1.3.1	4.4.2.7	A-20
Wellman-Lord FGD Process			
Thiosulfate/sulfate purge	4.2.2.1.3.2	not evaluated	A-10

(continued)

TABLE 3-83. (Continued)

Control Technique/Secondary Waste	PCTM Section Reference for Waste Characterization	PCTM Section Reference for Control Technique	Appendix
Coal Particulate Dry Collectors			
Coal Particulate	4.2.6	not evaluated	A-11, A-12
Water Pollution Control			
Phosam-W Process			
Acid gas	4.3.1.1.3.1	4.2.1.1	A-6, A-8, A-9, A-10
Chevron-WWT Process			
Acid gas	4.3.1.1.3.2	4.2.1.1	A-6, A-8, A-9, A-10
Activated Sludge			
Waste sludge	4.3.1.1.4.1	4.4.4.1.1	B-10, C
Activated Cargon Adsorption			
Regeneration off-gas	4.3.1.1.5.1	4.3.1.1.5.1	B-15
Incineration			
Flue gas	4.3.1.1.5.3	4.3.1.1.5.3	A-11, A-14, A-20, A-21
Cooling Tower			
Evaporation and drift	4.3.1.1.6.1	not evaluated	B-1, B-2, B-4, B-19
Blowdown	4.3.1.1.6.1	4.3.2	
Vapor Compression Evaporator			
Waste brine concentrate	4.3.1.1.6.2	not evaluated	C
Solid Waste Management			
Incineration			
Bottom ash	4.4.4.1.1	4.4.4.1.1	C

*Control techniques for these secondary waste streams are similar to those of other waste streams discussed in Sections 4.2, 4.3 and 4.4.

4. EVALUATION OF POLLUTION CONTROL TECHNOLOGIES

4.1 INTRODUCTION

At the present time, no commercial scale EDS plants are operating or are under construction. Thus, control technologies which would be applicable to the waste streams identified in Section 3 have not been demonstrated on EDS facilities, or on other direct liquefaction facilities. The control technologies which will be applied to EDS installations will likely be adaptations of those which are in use in such related industries as petroleum refining, coal cleaning, by-product coke, and coal-fired power plants. This section provides an evaluation of the applicability, performance and cost of control methods which may be adapted from other industries and from general pollution control practice. The section defines the limitations of and necessary modifications to these control methods for use in coal liquefaction plants, and discusses those controls which have actually been demonstrated in coal gasification/liquefaction plants. The information presented in this section provides users with background on a variety of alternative control options, and permits assessment of techniques for his specific situation, and assessment of how individual techniques might be combined for the case under examination.

Throughout this section, the flow rates of waste streams and effluents from control devices and control costs (except unit cost) presented all correspond to an EDS commercial plant processing 1,134 Mg/hr (30,000 tons per stream day) of "as received" Illinois No. 6 coal in the liquefaction area. This EDS plant produces about 9,580 m^3 (60,240 barrels) fuel oil equivalent per stream day of liquid products for the base case design, or 11,300 m^3 (71,080 barrels) fuel oil equivalent per stream day of liquid and gaseous products for the MFS case design.

4.1.1 Organization of Section

Section 4 is organized into three major subsections (2 digit heading) by waste media (air, liquid and solid). In each media section, the waste streams identified in Section 3 and secondary waste streams from other pollution controls are grouped into source types according to the waste characteristics. A source type is a grouping of those streams which require the same or similar types of controls. Evaluation of the applicable controls for each source type will be presented in a separate subsection.

The source type subsection (3 digit heading) begins with a list of the streams that fall into that source type. Where controls for a given source type consist of several control techniques connected in series, each step in the series is referred to as a control function. Applicable control techniques are grouped into control functions according to their purposes or pollutants removed. For example, bulk sulfur removal from acid gases is considered a control function, and the Claus process is a control technique under this control function. The relations among the pertinent control functions and their purposes are briefly discussed at the beginning of each source type subsection. Individual control techniques for each control function are then summarized in tables at the beginning of this subsection, briefly indicating the process operating principles, performance, reliability, secondary waste streams generated, costs, and advantages/disadvantages.

Each source type subsection is broken down according to each of the streams within that source type; each stream is discussed in a subsection with a 4-digit heading. These streams can be individual streams, as defined in Section 3, or can be combinations of these streams where appropriate (e.g., Section 4.2.1.1 discussions apply to a combination of the individual acid gas streams from various sources within the EDS plant). Each stream subsection is then further subdivided into subsections for each applicable control function where appropriate (5-digit heading; e.g., Section 4.2.1.1.1 covers bulk sulfur recovery for the combined acid gases). Finally, each function subsection is subdivided into 6-digit subsections for each applicable alternative individual control technique (e.g., Section 4.2.1.1.1.1 discusses the Claus process for bulk sulfur removal applied to the combined acid gas stream). It is under this 6-digit subsection that the detailed discussion of the individual control techniques is presented as applied to the specific stream. This discussion of individual control techniques addresses only the unique design/operating/performance features associated with their application to the specific waste streams in an EDS commercial plant. The EDS base case and MFS case designs are considered in this manual, because characteristics of emissions from these two specific Exxon designs can be estimated. Other EDS designs could result in emission streams with different characteristics and flow rates; however, the emission streams for the EDS base case and MFS case designs are considered to be representative, and the control techniques for these two designs are likely to be equally applicable to other EDS designs.

Detailed discussions of each individual control technique supporting the presentation in Section 4 are found in the separate Appendix volume.

For air and water pollution control technologies, the last subsection under each source type presents integrated control examples, for selected waste streams which require multiple control techniques in series. For solid waste management, integrated control examples are presented after the discussion of all the source types. The integrated control examples illustrate some typical methods by which specific techniques might be combined in a control system. The integrated control examples are selected on the basis of their consideration in designs by direct liquefaction process developers, and being representative of a broad range of control possibilities. The integrated control examples presented are by no means inclusive of all potentially applicable combinations. Performance, costs, and other key features of the integrated control examples are presented.

The cross-reference tables at the end of Section 3 (Tables 3-82 and 3-83) indicate the subsections within Section 4 (and the Appendix sections) in which the reader will find the control techniques applicable to each individual stream identified in Section 3.

4.1.2 Approach

The various control technologies applicable to the management of each of the waste streams identified in Section 3 are evaluated in Section 4. The PCTM first discusses control functions that might be involved in the treatment of waste streams under each source type, and then presents general information concerning the key features of various processes for each control function. These key features include: operating principles control efficiency for components removed, feed stream requirements/restrictions, potential for recovery of by-products, secondary waste streams generated, process reliability, and unit capital cost. Information on the key features is based on the application of control techniques in related industries. Following the presentation of general information on control techniques, the application of selected control techniques to individual EDS waste streams (or combinations of waste streams) is examined in detail, to illustrate how performance and cost estimates can be derived. Finally, integrated control examples for selected waste streams are presented to illustrate how specific control techniques might be combined. The information presented will allow the user to assess which techniques might be most useful for his use, and assemble his own systems from the breakdown of control functions.

Performance data for applicable control technologies have been obtained primarily from the open literature supplemented by vendor- and user-supplied data in some cases. The capabilities of various controls have not usually been assessed on a design-specific basis, but rather upon a generalized basis derived from test results and/or engineering studies of the subject technologies. In many cases, performance can only be estimated in terms of control of major constitutents (e.g., total sulfur) or gross parameters (e.g., BOD, COD) since information for removal efficiencies of specific substances is lacking. Further, even in those cases where substance-specific performance information exists for a control technology, accurate or complete characterization of the EDS waste stream being fed to the control device may be lacking. In the final analysis, the capabilities of controls on an EDS facility can only be accurately evaluated by testing at operating facilities or of smaller units from which data can be confidently extrapolated to commercial size. The performance estimates in this document are believed to reflect the best information currently available based on actual experience (on experimental facilities and in related conventional industries), and based on engineering analysis.

Detailed information on performance of control technologies, supporting the presentation in Section 4, is provided in the Appendix volume.

4.1.3 Costing Methodology

In order to compare controls for cost effectiveness and to estimate the impact of pollution control costs on overall plant costs, approximate capital and operating costs are generally presented in three ways: 1) cost per unit throughput; 2) absolute cost for an EDS commercial plant that processes 1,134 Mg/hr (30,000 tons per stream day) of "as received" Illinois No. 6 coal in the liquefaction area; and 3) cost as a percentage of the uncontrolled EDS base plant. The cost per unit throughput is often presented in both the general discussion of the control techniques and the discussion of application of control techniques to individual waste streams. In the general discussion, unit costs are sometimes given as a range of values, to represent the impacts of throughput rates (e.g., lower unit cost for higher throughput rate), variations in stream characteristics (e.g., lower unit cost for removing pollutant from a stream of higher pollutant concentration, if the unit cost is computed on the basis of amount of pollutant removed), variations in designs for a given technique (e.g., equipment quality-service factor tradeoff), and the design for how the individual technique is integrated into the integrated control system (e.g., residence time and other factors in the biox sections of an integrated control system). In the application to individual waste streams, unit costs are typically given as a single value, because a specific set of design and operating conditions has been selected to control the waste stream, and the specific throughput rate of the waste stream (or amount of pollutant removed) is used to compute the unit cost. As with the case of absolute cost and cost as a percentage of the uncontrolled EDS base plant, this single value for the unit cost does not imply a firm cost estimate, but merely reflects that a single set of numbers (versus ranges) has been used to estimate the cost.

The capital investment and annualized costs are based primarily on factored estimates of costs contained in non-proprietary published literature, normalized to a first quarter 1980 basis using generally accepted cost indices such as the Chemical Engineering (CE) plant cost index. Generally, a conservative approach was taken so that an overestimation of actual costs is more likely than an underestimation. For example, when cost data for a control technique were scattered, the approach was to utilize the data points associated with the higher costs to develop cost curves for use in this manual. This is because there is a tradeoff between equipment quality and service factor. Selection of the higher cost data points as the cost basis implies that higher quality equipment are to be used, to insure less troublesome operation and higher service factor. It is also recognized that total costs are strongly influenced by oversizing and/or redundance designed into systems in order to compensate for the uncertainties in reliability or performance levels. However, no attempt has been made to include redundancy in the PCTM cost, although some redundancy is built into many of the original literature or vendor capital cost bases for the various controls. Detailed information on the costs of individual control processes is provided in the Appendix volume.

To the extent possible, the same methodology was used to develop capital and total annualized cost estimates for pollution controls and for the uncontrolled base plants (Section 2.3). Total capital investment costs (TCI) presented in the PCTM include:

1) purchased and delivered equipment costs;
2) labor and materials costs to install equipment;
3) indirect installation charges, such as
    - engineering and construction costs,
    - contractor fees, and
    - project and process contingency reserves;
4) interest expenses on capital spent prior to start of production (interest during construction)
5) working capital.

The first four items add up to the total depreciable investment (TDI). Start-up costs are not included and working capital is not depreciable. Total capital investment estimates for the uncontrolled EDS plant include working capital but not land costs. Total capital investment estimates for pollution control technologies, however, do not include non-depreciable items such as working capital and land costs, because working capital for individual control techniques is generally a small item and difficult to estimate. Thus, in the context of the EDS PCTM, the total capital investment (TCI) and the total depreciable investment (TDI) costs are the same for pollution control technologies.

A variety of methods can be used to estimate the above cost items, although most methodologies utilize a factored approach. In a factored approach, the cost of purchased and delivered equipment is obtained from vendor quotes or estimated from previous projects using similar equipment. The remaining cost items are then estimated as a "factor" times the purchased and delivered equipment costs or other subsequently derived subtotal.

For cost estimates developed for this PCTM, the major source of cost information was the open literature, although some vendor quotes were used. In general, literature cost information is not reported as delivered equipment costs: some data are published as installed equipment costs (purchased equipment plus direct installation costs), some include one or more of the indirect installation charges (listed previously), and some are TDI estimates. It was beyond the scope of the PCTM to develop the detailed engineering designs necessary for cost estimation at the "firm" (± 5 to 10%) level. The accuracy of cost data taken from published sources was influenced by the reliability of the data source, the cost methodology, and the degree of similarity of the streams. Also, the accuracy of the estimates and the components in the cost estimates (e.g., contingency) are not always provided in the reference. Thus, extrapolation of these costs to the stream being treated in the PCTM also introduced uncertainties. Although the accuracy of the cost estimates vary, most are believed to be within ± 50% for the 1980 base year. Uncertainties in cost estimates due to uncertainties in waste stream characteristics, however, are not included in this ± 50% figure.

In order to provide consistency in the various capital cost estimates required for the PCTM, a capital cost methodology was developed. The methodology (and cost factors) used are summarized in Table 4-1. Most cost data obtained from the literature were installed equipment costs (IEC), which represent the total direct capital costs. This cost includes principal equipment items together with installation of these equipment items, piping and instrumentation, minor steelwork such as ladder platforms, supports (not major structures), concrete foundations and substructures, electrical, insulation, buildings housing equipment and paint. For cost data available as equipment cost, a field materials and labor (M&L) factor was applied to relate the purchased equipment cost to the installed equipment cost. The field materials and labor factor usually ranged from 1.8 to 2.6, depending on the type of equipment to be installed. In most cases, the installed equipment cost (IEC) or the purchased equipment cost (C) was available for equipment at a different operating capacity. The general approach in correlating costs for equipment at different capacities

TABLE 4-1. CAPITAL COST ESTIMATING METHOD

Installed Equipment Costs (IEC)

  Principal equipment items, installation of these items, piping, instrumentation, foundations and substructures, electrical, insulation, buildings housing equipment, and paint.

Indirect Installation Costs (IIC)

- Engineering and Construction (25% of IEC)
- Fees (3% of IEC)
- Contingency (20% of IEC)

Total Plant or Process Costs (TPC = IEC + IIC)

Interest During Construction (IDC = 22.6% of TPC)

Total Depreciable Investment (TDI = TPC + IDC)

Working Capital*(WC = 60 days inventory of raw materials, catalysts and chemicals)

Total Capital Investment*(TCI = TDI + WC) was to utilize the relationship:

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*Total capital investment estimates for pollution control technologies do not include non-depreciable items such as working capital and land costs; i.e., it is assumed that WC is small and TCI approximately equals TDI.

Cost basis: 1980 dollars.

$$C_B = C_A (Q_B/Q_A)^n$$

where: C_A, C_B = cost of equipment at capacity A or capacity B (C_A is known)
Q_A, Q_B = operating capacity of A or B (Q_A and Q_B are known)
n = size exponent factor

The operating capacity factor (Q) in the above equation can refer to a volumetric flow rate, quantity of sulfur removed or quantity of coal processed, for example. The size exponent factor (n) normally ranges from 0.4 to 0.9, depending upon the type of equipment or process being evaluated. If specific size exponents were not available in the literature, a value of 0.6 was used (the "six-tenths rule" of equipment cost estimation). Alternatively, if sufficient published cost data were available, cost curves were developed.

As indicated in Table 4-1, indirect installation charges were estimated as 48% of the IEC. Adding the indirects to the IEC gave the total plant (or process) costs. Interest during construction (IDC) was estimated as 22.6% of the total plant costs, based on 12% per year for an average of 1.88 years. The TDI is the sum of the total process cost (TPC) and the interest during construction (IDC).

Annualized costs consist of annual operating expenses plus annualized capital-related charges. Annual operating expenses include costs for labor (operating, supervision, and maintenance), raw materials, chemicals, catalysts, utilities (steam, electricity, cooling water, etc.), and overhead. Capital-related charges include interest on working capital, local taxes, insurance, depreciation, income taxes, and return on investment. The unit costs or factors used to estimate total annualized costs in this PCTM are listed in Table 4-2. All of the terms listed in Table 4-2, except capital recovery, are expressed in first year costs, i.e., in constant 1980 dollars. The capital recovery term, however, is a levelized value calculated using standard present worth and levelized cost procedures and the economic assumptions listed in Table 4-3.

TABLE 4-2. UNIT COSTS AND FACTORS FOR ANNUALIZED COST ESTIMATES

Operating Labor ($11/hr)
Supervision (15% of operating labor)
Maintenance (2% of total depreciable investment)
Maintenance Supervision (5% of maintenance)
Illinois No. 6 Coal ($1.14/GJ)
Raw Water (\$0.031/m^3)
Utilities

  - Steam ($5.73/Mg)
  - Electricity ($0.033/kW-hr)
  - Fuel Gas ($4.43/GJ)
  - Cooling Water (\$0.08/m^3)

Chemicals and Catalysts (representative early 1980 costs)

By-product Credit

  - Ammonia ($140/Mg)
  - Phenol ($6.05/l)
  - Sulfur (0)

Overhead Charges

  - Plant overhead (50% of operating labor)
  - General and administrative overhead (15% of operating labor and maintenance)

Laboratory Charges (5% of operating labor)

Capital Related Charges (17.2% of total depreciable investment)

TOTAL ANNUALIZED COSTS (summation of above items)

TABLE 4-3. ASSUMPTIONS USED TO CALCULATE CAPITAL RECOVERY FACTOR

Financing basis:	100% equity
Desired after tax return on investment:	12% of total depreciable investment
Income tax rate:	48% of taxable income
Economic facility life:	20 years
Facility life for depreciation purposes:	16 years
Depreciation method:	sum-of-the-years-digit
Investment tax credit:	20% of total depreciable investment
Interest rate	12% per year
Property taxes and insurance	3% of total depreciable investment

4.2 AIR POLLUTION CONTROL TECHNOLOGIES

Uncontrolled gaseous emissions from the EDS commercial plant have been identified and described in Section 3 of this manual. Characteristics of these streams and additional gaseous waste streams generated by pollution control processes from other media are summarized in Tables 4-4 and 4-5 for the base case and MFS case designs, respectively. The waste streams presented in these tables are regrouped into six source types according to the major types of potential pollutants which they contain:

- Source Type 1 - acid gases and other reduced sulfur/nitrogen, organic-laden gases
- Source Type 2 - combustion gases
- Source Type 3 - organic and CO containing waste gases
- Source Type 4 - fugitive dusts from material storage
- Source Type 5 - fugitive organic emissions
- Source Type 6 - fugitive particulates from material conveying and processing.

The grouping of waste streams into source types facilitates the discussion of air pollution control technologies, because similar controls are generally applicable to waste streams belonging to the same source type. The characteristics of waste streams presented in Tables 4-4 and 4-5 are those that are most relevant to the selection and evaluation of pollution control technologies. The flow rates and compositions are based upon an EDS commercial plant processing 1,134 Mg/hr (30,000 tons per stream day) of "as received" Illinois No. 6 coal in the liquefaction area (compositions of the design Illinois No. 6 coal are given in Table 2-1). These flow rates and compositions might, of course, vary in practice due to differences in plant design, coal type or plant operating conditions, and due to uncertainties in estimation.

In the ensuing subsections, air pollution control technologies which may be applicable to the gaseous waste streams listed in these tables are identified. Some of these gaseous waste streams would be combined for treatment

TABLE 4-4. ESTIMATED CHARACTERISTICS OF GASEOUS WASTE STREAMS ACCORDING TO SOURCE TYPE FOR EDS COMMERCIAL PLANT (ILLINOIS COAL BASE CASE)

Source Type/Stream	Stream Flow Rate[†] kmol/hr	Stream Characteristics
Source Type 1 - Acid Gases and Other Reduced Sulfur/Nitrogen, Organic-Laden Gases		
Stream 508 - acid gas from DEA regenerator	1,214	Contains 64.2% H_2S, 371 ppmv COS, and 3.9% NH_3
Stream 501 - acid gas from sour water stripper/ ammonia recovery[‡]	263	Contains 54.0% H_2S and 0.7% NH_3
Stream 304 - Flexicoking gasifier/heater sour fuel gas	54,781	Contains 0.42% H_2S, 108 ppmv COS, and 11.6% CO_2
Stream 102 - slurry drier vent gas	532	Contains 30 ppmv H_2S and 4.1% hydrocarbons
Stream 153 - vacuum fractionator off-gas	69	Contains 5% H_2S and 47.5% hydrocarbons
Stream 803 - transient waste gas from liquefaction reactor	3,280	Contains 2.5% H_2S and 45.1% hydrocarbons
Stream 801 - transient waste gas from Flexicoking	671	Contains 1.0% H_2S and 40.9% hydrocarbons
Source Type 2 - Combustion Gases		
Stream 707a - flue gas from power generation system	36,450	Contains 2850 ppmv SO_2, 359 ppmv NO_x and 6760 mg/m^3 particulate matter
Stream 701a - flue gas from steam generation system	42,560	Contains 925 ppmv SO_2, 152-187 ppmv NO_x, and 2140 mg/m^3 particulate matter
Stream 107 - flue gas from liquefaction slurry preheat furnace	34,535	Contains 41 ppmv SO_2, 61-117 ppmv NO_x, and 4.9-14.6 mg/m^3 particulate matter
Stream 203 - flue gas from solvent hydrogenator fuel preheat furnaces	3,864	Contains 41 ppmv SO_2, 61-117 ppmv NO_x, and 4.9-14.6 mg/m^3 particulate matter
Stream 434 - flue gas from hydrogen plant reformer furnaces	58,022	Contains 41 ppmv SO_2, 61-117 ppmv NO_x, and 4.9-14.6 mg/m^3 particulate matter
Source Type 3 - Organic and CO Containing Waste Gases		
Stream 426 - vent gas from CO_2 removal by Catacarb process	6,100	Contains 361 ppmv CH_4, 1.9% H_2, and 157 ppmv CO
Stream 438 - hydrogen plant deaerator vent	6.5	Consists primarily of steam with a small quantity of CO_2 and traces of hydrocarbons
Stream 446 - regeneration/decommissioning off-gas from reformer catalyst	6,800	Consists primarily of steam and nitrogen with small amounts of CO and particulate matter
Stream 448 - decommissioning off-gas from methanation catalyst	5,000	Consists primarily of steam and nitrogen with small amounts of particulate matter
Source Type 4 - Fugitive Dust from Material Storage		
Stream 011 - fugitive dust from coal pile	N/A[*]	Emits an average of 51 kg/hr of particulate matter
Source Type 5 - Fugitive Organic Emissions		
Stream 751 - evaporative emissions from product and by-product storage	0.02	Contains mostly aliphatic hydrocarbons, benzenes and substituted benzenes, and phenols
Fugitive organic emissions from pump and compressor seals, valves, flanges, pressure relief valves, and drains	1.5	Emits 153 kg/hr of organics
Source Type 6 - Fugitive Particulates from Material Conveying and Processing		
Stream 013 - fugitive particulates from coal handling and crushing	N/A	Emits 161-2,548 kg/hr of particulate matter

[*] N/A - not applicable

[†] **For EDS plant producing 9,580 m^3 (60,240 barrels) fuel all equivalent products per stream day.**

[‡] **Secondary discharge stream**

TABLE 4-5. ESTIMATED CHARACTERISTICS OF GASEOUS WASTE STREAMS ACCORDING TO SOURCE TYPE FOR EDS COMMERCIAL PLANT (ILLINOIS COAL MFS CASE)

Source Type/Stream	Stream Flow Rate,[†] kmol/hr	Stream Characteristics
Source Type 1 - Acid Gases and Other Reduced Sulfur/Nitrogen, Organic-Laden Gases		
Stream 508 - acid gas from DEA regenerator	1,214	Contains 64.1% H_2S, 373 ppmv COS, and 5.2% NH_3
Stream 501 - acid gas from sour water stripper/ammonia recovery[‡]	243	Contains 56.3% H_2S and 0.7% NH_3
Stream 440 - flash gas from partial oxidation unit	254	Contains 4.4% H_2S and 1180 ppmv COS
Stream 428 - acid gas from acid gas removal unit in hydrogen purification	13,304	Contains 1.0% H_2S, 38 ppmv COS, and 669 ppmv NH_3
Stream 304 - Flexicoking gasifier/heater sour fuel gas	27,225	Contains 0.44% H_2S, 92 ppmv COS, and 11.5% CO_2
Stream 102 - slurry drier vent gas	532	Contains 30 ppmv H_2S and 4.1% hydrocarbons
Streams 153/156 - vacuum fractionator off-gas	69	Contains 5% H_2S and 47.5% hydrocarbons
Stream 803 - transient waste gas from liquefaction reactor	3,280	Contains 2.5% H_2S and 45.1% hydrocarbons
Stream 801 - transient waste gas from Flexicoking	511	Contains 1.5% H_2S and 39.7% hydrocarbons
Stream 802 - transient waste gas from partial oxidation units	12,121	Contains 0.3% H_2S, 250 ppmv COS, and 0.2% hydrocarbons
Source Type 2 - Combustion Gases		
Stream 707a - flue gas from power generation system	26,110	Contains 2850 ppmv SO_2, 359 ppmv NO_x, and 6760 mg/m^3 particulate matter
Stream 701a - flue gas from steam generation system	69,150	Contains 1470 ppmv SO_2, 210-236 ppmv NO_x, and 3450 mg/m^3 particulate matter
Stream 107 - flue gas from liquefaction slurry preheat furnace	34,535	Contains 41 ppmv SO_2, 61-117 ppmv NO_x, and 4.9-14.6 mg/m^3 particulate matter
Stream 161 - flue gas from partial oxidation feed vacuum fractionator preheat furnace	425	Contains 41 ppmv SO_2, 61-117 ppmv NO_x, and 4.9-14.6 mg/m^3 particulate matter
Stream 203 - flue gas from solvent hydrogenation fuel preheat furnaces	3,864	Contains 41 ppmv SO_2, 61-117 ppmv NO_x, and 4.9-14.6 mg/m^3 particulate matter
Streams 449/450 - regeneration/decommissioning off-gas from high and low temperature shift catalysts	22,255	Consists primarily of steam with 6% N_2, 1% SO_2, and 0.5% CO_2
Source Type 3 - Organic and CO Containing Waste Gases		
Stream 438 - hydrogen plant deaerator vent	6.2	Consists primarily of steam with a small quantity of CO_2 and traces of hydrocarbons
Source Type 4 - Fugitive Dusts from Material Storage		
Stream 011 - fugitive dust from coal pile	N/A*	Emits an average of 52 kg/hr of particulate matter
Source Type 5 - Fugitive Organic Emissions		
Stream 751 - evaporative emissions from product and by-product storage	0.02	Contains mostly aliphatic hydrocarbons, benzenes and substituted benzenes, and phenols
Fugitive organic emissions from pump and compressor seals, valves, flanges, pressure relief valves, and drains	1.5	Emits 153 kg/hr of organics
Source Type 6 - Fugitive Particulates from Material Conveying and Processing		
Stream 015 - fugitive particulates from coal handling and crushing	N/A	Emits 165-2,623 kg/hr of particulate matter

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* N/A - not applicable

[†] For EDS plant producing 11,300 m^3 (71,080 barrels) fuel oil equivalent products per stream day.

[‡] Secondary discharge stream

rather than handled separately; for these streams, stream combinations are identified, and controls are presented for the combined stream rather than for each individual stream. Further, some of these waste streams would require treatment by a series of control technologies, rather than by just a single technique; for these streams, the individual control techniques are broken down according to the control function that they fulfill within the total control system; and, in addition to discussions of individual techniques, example approaches to the integrated control of selected waste streams or stream combinations are provided. In the discussions which follow, emphasis is placed on streams unique to the EDS process, rather than on those streams which are exactly identical to streams found in conventional industries (e.g., fugitive dust from coal piles, flue gas from coal-fired boilers).

In Tables 4-4 and 4-5, the acid gas from sour water stripper/ammonia recovery (Stream 501) is a secondary waste stream generated from water pollution control. Secondary gaseous waste streams -- generated from air pollution control (hence not cross-media) -- will be introduced as the control technique causing their generation is discussed. Controls for secondary waste streams will sometimes be handled as part of a combined stream (as will be the case for Stream 501), and sometimes be handled separately but often following the discussion of the control technique causing their generation.

4.2.1 Source Type 1 - Acid Gases and Other Reduced Sulfur/Nitrogen, Organic-Laden Gases

There are seven gaseous waste streams under Source Type 1 for the base case design, and eleven waste streams under Source Type 1 for the MFS case design. All seven gaseous waste streams for the base case design are also present for the MFS case design. These waste streams are:

- Stream 508 - acid gas from DEA regenerator
- Stream 501 - acid gas from sour water stripper/ammonia recovery
- Stream 440 - flash gas from partial oxidation unit (MFS case only)
- Stream 428 - acid gas from acid gas removal unit in hydrogen purification (MFS case only)
- Stream 304 - Flexicoking gasifier/heater sour fuel gas
- Stream 102 - slurry drier vent gas
- Stream 153 - vacuum fractionator off-gas
- Stream 156 - partial oxidation feed vacuum fractionator off-gas (MFS case only)
- Stream 803 - transient waste gas from liquefaction reactor
- Stream 801 - transient waste gas from Flexicoking
- Stream 802 - transient waste gas from partial oxidation unit (MFS case only).

All gaseous waste streams are generated from the uncontrolled EDS base plant, with the exception of Stream 501, which is generated from treatment of the sour water containing H_2S and NH_3.

Approaches to treatment of gaseous waste streams under Source Type 1 may involve up to three control functions:

1) bulk sulfur removal
2) residual sulfur removal
3) incineration.

Bulk sulfur removal, as the name implies, removes most of the sulfur compounds (e.g., 95%) present in the waste stream. Residual sulfur removal is designed to reduce the last remaining sulfur compounds in the off-gas from bulk sulfur removal down to very low levels, employing processes that can operate in this low concentration region more effectively than bulk sulfur removal processes, based on either technical or economical considerations. Incineration oxidizes reduced sulfur species in the gas stream prior to sulfur removal by SO_2 control techniques. It is also used to oxidize reduced sulfur species remaining after bulk or residual sulfur removal, if necessary, for control of odor in emission streams.

Depending on the characteristics of the waste stream and the specific control technique(s) employed, it may not be necessary to utilize all three control functions in any treatment approach. For example, there may not be a need for incineration following residual sulfur removal. For the EDS process, pretreatment or enrichment of acid gas streams (Streams 508 and 501) prior to bulk sulfur removal is not necessary because of the concentration levels of H_2S in these streams. Also, there are some options in the sequential order of applying these control functions. For example, incineration may precede bulk sulfur removal (e.g., incineration followed by flue gas desulfurization) or may follow residual sulfur removal (e.g., incineration of the off-gas from sulfur plant tail gas treatment).

Partial or essentially complete control of reduced nitrogen compounds (NH_3 and HCN, if present), CO, and organics can be realized either as an integral part of the sulfur control approach or through a separate add-on step such as incineration.

Bulk Sulfur Removal

To date only three processes have seen any significant commercial application for the bulk removal of H_2S from acid or fuel gases, namely the Claus, Stretford, and Giammarco-Vetrocoke processes. Although a few other bulk sulfur removal processes are available or have been proposed, these three are the only processes examined here for a variety of reasons, including: expressed interest to date by synfuels developers, advanced commercial status and an extended operating history, applicability to a wide range of sulfur concentration levels in the acid gases, and availability of performance and cost data. In addition to these processes, alternatives involving incineration followed by SO_2 removal may be applicable for treatment of sour fuel gas. In conventional applications, Stretford units have historically been favored over Claus units due to economic reasons for feeds containing only a few percent H_2S. Claus plants are often preferred for feeds containing over 25% H_2S. The Giammarco-Vetrocoke process is generally applicable to feed streams with H_2S concentrations of up to 1.5%. Table 4-6 summarizes the key features of these bulk sulfur removal processes. These three bulk sulfur removal processes are described below. Further details on the Claus and Stretford processes and supporting performance and cost data for these processes can be found in Appendices A-6 and A-7 of the Appendix volume.

The Claus process is a dry, high temperature process in which H_2S is catalytically reacted with SO_2 to form elemental sulfur. There are two common versions of the process: "straight through" and "split flow". In the "straight through" mode the total acid gas stream (generally with all the sulfur in the form of H_2S or other reduced sulfur species) is sent to the combustion chamber, and sufficient air is added to oxidize one-third of the H_2S to SO_2, in order for the Claus reaction to proceed. The "split flow" mode is often used when H_2S-levels in the feed gas are below 25% by volume. The acid gas is split into two streams and one-third of the input acid gas is combusted in a reaction furnace to form SO_2. Heat is recovered from the combusted gas before it is recombined with the other two-thirds of the feed. The combined stream then enters a series of converter stages where elemental sulfur is produced and recovered. There are also special considerations for acid gas feeds containing high levels of ammonia. Residual ammonia would form salts with sulfur

TABLE 4-6. KEY FEATURES OF BULK SULFUR REMOVAL PROCESSES

	Claus	Stretford	Giamarco-Vetrocoke (G-V)	Incineration/SO_2 Removal
Principle of Operation	Catalytic oxidation of H_2S and SO_2 to elemental sulfur.	Liquid phase oxidation of H_2S to elemental sulfur in an alkaline solution of metavanadate and anthraquinone disulfonic acid.	Liquid phase oxidation of H_2S to elemental sulfur in sodium carbonate and arsenate/arsenite solution.	Oxidation of reduced sulfur and organics, followed by SO_2 removal using either regenerative or throwaway FGD technologies.
Components Removed	H_2S, COS, RSH, VOC, NH_3, and HCN.	H_2S, HCN, and CH_3SH.	H_2S, COS, and CS_2.	H_2S, COS, RSH, VOC, and CO.
Efficiency	Over 95% total S, other combustibles partially destroyed.	As low as 1 ppmv H_2S but no removal of non H_2S sulfur.	99.99% H_2S removal.	As low as 100 ppmv VOC in incinerated gas and up to 99% total sulfur removal.
Feed Stream Requirements/ Restrictions	Streams containing H_2S levels much below 10% require enrichment prior to processing. Organics cause combustion control problems and "grey" sulfur.	High HCN loading should be reduced prior to processing to prevent excessive solution purge.	Maximum 1.5% H_2S in feed.	In principle, gases with any level of H_2S or sulfur compounds could be incinerated and subsequently treated via FGD. Other components cause no problem.
By-Products	Elemental sulfur.	Elemental sulfur.	Elemental sulfur which may require arsenic removal.	Either $CaSO_4$, concentrated SO_2, or throwaway lime sludges are generated by FGD units.
Secondary Waste Streams	Spent catalyst and catalyst regeneration decommissioning offgas.	Oxidizer vent gas and purge solution.	Oxidizer vent gas and arsenate/arsenite wash water.	Some condensate and scrubber sludge.
Reliability/ Limitations	NH_3 and HCs may cause catalyst plugging and variable sulfur recovery.	Process does not remove COS, RSH, or organics, HCN forms nonregenerable salts in scrubbing solution.	Hazardous nature of arsenic solution may cause handling and safety problems.	FGDs systems have varying degrees of reliability and generally have lower on stream factors than process units.
Effects of High CO_2 in Feed	Can adversely affect sulfur removal ability of the process.	High CO_2 concentrations will decrease absorption efficiency by lowering solution alkalinity. Increased absorber tower height and addition of caustic are required.	Little or no effect.	No effect except to increase energy requirement for incineration if insufficient combustibles are present in feed gas.
Capital Costs	\$25 to \$180 x 10^3 per Mg sulfur/day capacity depending on both total flow and sulfur content.	\$110 to \$270 x 10^3 per Mg sulfur/day capacity, depending primarily upon total flow.	No cost data available.	\$700 to \$1700 x 10^3 per Mg sulfur/day capacity depending upon total flow and degree of sulfur removal.
General Comments	Applicable only to acid gases from selective AGR system. Hydrocarbon removal from feeds may be necessary.	1 ppmv H_2S in tail gas is possible, however higher limits are proposed when high levels of other reduced sulfur species are present in tail gas.	Limited data available. Hazardous nature of arsenic solution makes application unlikely in large U.S. facilities.	FGD process has usually been applied to combustion flue gases containing less than 5000 ppmv SO_2 and achieving about 90% control. Performance and cost data for higher SO_2 feeds achieving 99% control are limited.

compounds and carbon dioxide which cause plugging and catalyst deactivation problems downstream. Therefore, the ammonia-bearing contributor to the total acid gas stream is often fed to a separate compartment of the combustion chamber, so that all the ammonia can be entirely combusted to primarily nitrogen and water. The other acid gas streams might be fed to the second compartment of the combustion chamber, along with additional air, if necessary, so that overall one-third of the H_2S in the total acid gas feed is combusted to form SO_2.

Regardless of the Claus mode, the number of stages determines sulfur removal efficiency; Claus units of 3-stage design can achieve overall removal efficiencies of over 95%. To further reduce tail gas sulfur emissions, the Claus sometimes uses four catalytic stages for up to 97% sulfur recovery, at which point increased plant efficiency by this technique is limited by thermodynamic considerations. Gaseous sulfur species distribution in Claus tail gas in high CO_2 applications is approximately 60% H_2S, 30% SO_2, 9% COS, and 1% CS_2. Elemental sulfur as both vapor and entrained mist can contribute 20-50% to the total sulfur in Claus tail gases, depending primarily on the level of H_2S in the Claus feed and the effectiveness of mist eliminators. The relative contribution of elemental sulfur to total sulfur in Claus tail gas generally increases as H_2S content of Claus feed gas decreases.

In the "straight through" mode of Claus operation, organics, HCN, and NH_3 in the feed are largely converted to carbon dioxide, water vapor, and elemental nitrogen. Such components are not ordinarily of concern to process operation unless levels exceed perhaps 1% each. Organics make control of combustion stoichiometry and temperature more difficult and can lead to a by-product sulfur containing elemental carbon (black sulfur). HCN at high levels causes corrosion throughout the process while NH_3 can form deposits which plug/deactivate Claus catalysts. The organics problem is usually solved by limiting their content in the Claus feed. HCN at high levels can be destroyed (converted to NH_3) prior to entering the Claus furnace using Claus or shift type catalysts under reducing conditions. This should not be a problem with the EDS process. Ammonia at high levels require either bulk removal prior to Claus, special design to control temperatures and minimize deposition of ammonia salts, or split flow mode of operation. Because of the concentration levels of H_2S in the acid gas streams, the "straight through" mode would possibly be preferred in Claus plants designed for the EDS process.

The Claus process produces spent bauxite or alumina catalyst and catalyst regeneration off-gases where catalyst regeneration is used. The Claus catalyst has an estimated life of at least two to three years. Regeneration of catalyst is performed intermittently at a few facilities when the efficiency of the process drops below acceptable levels; no data regarding regeneration frequency, duration or off-gas characteristics are available. Further details concerning the Claus process are given in Appendix A-6.

The Stretford process is suited for the treatment of sour fuel gas and vent gas streams containing only a few percent H_2S in the EDS commercial plant. The original Stretford process (as developed by the British Gas Corporation) is a liquid-phase oxidation process using an aqueous carbonate/bicarbonate solution containing sodium vanadate and anthraquinone disulfonic acid (ADA) in which H_2S is both absorbed and converted to sulfur. The H_2S is absorbed by the carbonate/bicarbonate solution in either a packed tower (or contacted in a venturi scrubber) and then oxidized to sulfur by the sodium vanadate. The resulting reduced vanadium is then oxidized by the ADA Solution. ADA is in turn regenerated by air in separate oxidizer tanks where elemental sulfur is removed as a froth. A continuous solution purge is required to remove the buildup of sodium thiosulfate and sodium thiocyanate. Concentrations of H_2S in the purified gas stream from the Stretford process can typically be reduced to less than 10 ppmv, and sometimes to less than 1 ppmv. Carbonyl sulfide, carbon disulfide and lower hydrocarbons (e.g., containing 5 carbon atoms or less) are not removed by the Stretford solution and essentially appear in the purified gas stream. Mercaptans, however, are split between the purified gas stream and the oxidizer vent gas. Higher hydrocarbons as well as ammonia are almost completely removed and will all be present in the oxidizer vent gas. HCN is also completely removed, converted to SCN^-, and will appear in the solution purge.

Recently, modifications of the original Stretford process have been developed. One version of the Stretford process has been used at the SASOL Lurgi coal gasification complex in South Africa. At SASOL, severe plugging problems have occurred in the Stretford towers which apparently relate to the high CO_2 levels in the Stretford feed compared to feeds in other services. Preliminary information indicates that sulfur deposition is primarily responsible. SASOL has modified the original Stretford unit, presumably substituting a different absorbent but utilizing the bulk of the existing equipment.

The Stretford process generates two waste streams, the oxidizer vent gas and the purge solution. The purge solution may be treated via the reductive incineration process where the sulfur content of the purge is recovered as H_2S for recycle to the absorber and sodium and vanadium salts are recovered for reuse. The oxidizer vent gas is expected to consist primarily of air, carbon dioxide and water vapor. Further details concerning the Stretford process are provided in Appendix A-7.

The Giammarco-Vetrocoke H_2S removal process is a liquid phase oxidation process using an absorbent solution of alkali arsenates/arsenites in which hydrogen sulfide is both absorbed and converted to elemental sulfur. Sodium carbonate is the alkali usually applied for removal of sulfur because of its relatively low cost. The Giammarco-Vetrocoke process is applicable to gas streams containing up to 1.5% hydrogen sulfide, and can reduce hydrogen sulfide levels to 0.5 ppmv or less (43). The successive reactions occurring are (42):

$$Na_3\ As\ O_3 + 3H_2S = Na_3\ AsS_3 + 3\ H_2O \quad (4\text{-}1)$$

$$Na_3\ AsS_3 + 3\ Na_3AsO_4 = 3\ Na_3AsO_3S + Na_3\ AsO_3 \quad (4\text{-}2)$$

$$Na_3\ As\ O_3S = Na_3\ AsO_3 + S \quad (4\text{-}3)$$

$$Na_3\ AsO_3 + 1/2\ O_2 = Na_3\ AsO_4 \quad (4\text{-}4)$$

The hydrogen sulfide is absorbed at pressures from 0.1 to 7.5 MPa by countercurrent absorption, forming sodium thioarsenite according to reaction (4-1). The resulting thioarsenite is converted to monothioarsenate by reaction with arsenate (reaction (4-2)). This reaction occurs in the absorber and in the subsequent oxidizing column. Rich solution from the absorber is passed to an atmospheric pressure, air-blown oxidizing column. Under the oxidizing conditions, the monothioarsenate decomposes to arsenite and elemental sulfur (reaction (4-3)). Product sulfur is recovered by froth flotation, filtered and washed. The oxidizing reaction (4-4) re-establishes the arsenite/arsenate balance of the Vetrocoke solution by oxidizing some arsenite to arsenate. Based upon limited available data, the only waste streams generated by this process are wash water from the sulfur washing operation and oxidizer vent gas. Characterization data are not available for these streams although the wash water will contain arsenate/arsenite absorber solution.

One other approach to bulk sulfur control should be mentioned. In principle acid gases can be directly incinerated to convert all organics, CO, and reduced sulfur and nitrogen species to primarily CO_2, H_2O, N_2 and SO_2. SO_2 removal from the incinerated gas could then be accomplished using any one of a number of available FGD processes (see Section 4.2.2). Generally, such an approach is unattractive for several reasons. Throwaway FGD creates large solid waste disposal problems and recovery type FGD systems often feature Claus or Claus type (e.g., the Allied process) processes for elemental sulfur recovery; there is little to be gained in these approaches over direct use of the Claus process. FGD systems are also not demonstrated for gases containing over about 0.5% SO_2, and although there appear to be no inherent technical limitations prohibiting designs for much higher SO_2 levels, such systems are expected to be several times more costly than Claus or Stretford plants applied directly to crude acid gases. However, SO_2 removal by FGD processes might be competitive for the treatment of incinerated sour fuel gas, as compared to the alternative of removing sulfur from the fuel gas prior to combustion. Also, incineration followed by throwaway FGD processes might be attractive for the control of intermittent waste streams such as transient waste gases.

Residual Sulfur Removal

The Claus process is efficient in recovering up to 97% of the sulfur feed by using four catalytic stages for conversion of hydrogen sulfide in acid gas streams to elemental sulfur. Additional catalytic stages, however, would not result in increased sulfur recovery by the Claus technique due to thermodynamic limitations. The need for further reduction in emissions of sulfur compounds, therefore, has resulted in the development of residual sulfur removal processes capable of removing the residual traces of sulfur compounds present in the tail gases of bulk sulfur removal processes. A number of processes are now commercially available for treatment of sulfur plant tail gases or other waste gases containing low levels of reduced sulfur species. Table 4-7 summarizes the key features of the most prominent of these processes. The processes can be categorized in three generic types:

1) Conversion of sulfur species to H_2S followed by its removal using absorption/reaction - these include processes such as the Beavon Sulfur Removal Process (BSRP), the Shell Off-Gas Treating (SCOT) process, the Cleanair process, and the BSR/Selectox process.
2) Conversion of sulfur species to sulfur dioxide (SO_2) by incineration, followed by SO_2 removal - such as the Wellman-Lord, IFP-2, and Chiyoda Thoroughbred 121 processes.

TABLE 4-7. KEY FEATURES OF RESIDUAL SULFUR REMOVAL PROCESSES

	Beavon	SCOT	Incineration Plus SO_2 Removal (FGD)	Sulfreen
Principle of Operation	Catalytic reduction of sulfur compounds to H_2S, followed by integrated Stretford process.	Sulfur species are catalytically reduced to H_2S; H_2S is scrubbed in a regenerable amine system.	Incineration (an on-site boiler or separate incinerator) followed by SO_2 removal (e.g., Wellman-Lord).	Gas phase continuation of Claus reaction at a low temperature.
Components Removed	H_2S, COS, CS_2, and SO_2.	H_2S, COS, and CS_2, SO_2.	SO_2, also removes HCs, CH_3SH, NH_3, and HCN.	H_2S, SO_2, COS, and CS_2.
Efficiency	Over 99.9% total sulfur removal in combination with the Claus plant or can attain equivalent of 50 ppm total sulfur in tail gas (not including reducing gases).	Over 99.9% total sulfur removal in combination with the Claus plant or can attain equivalent of 250 ppm total sulfur in tail gas (will vary depending on CO_2 and H_2S concentration in specific applications).	Up to 99% total sulfur removal from Claus tail gas or 50 ppm SO_2 in tail gas and complete removal of other compounds.	Up to 99% sulfur removal in combination with Claus plant. Can expect a typical total sulfur level of 2500 ppm in tail gas.
Feed Stream Requirements/ Restrictions	None.	None.	None.	Optimum performance requires $H_2S:SO_2$ ration of 2:1.
By-Products	Elemental sulfur.	Concentrated H_2S.	Sulfur or sulfuric acid from the Wellman-Lord recovery FGD process.	Elemental liquid sulfur.
Secondary Waste Streams	Sour condensate, oxidizer vent gas, solution purge, and spent catalyst.	Sour condensate and spent catalyst.	Sour condensate and solution purge.	Spent catalyst.
Reliability/ Limitations	Has only been applied to Claus process tail gases.	Requires further treatment and/or recycle to Claus.	Solid wastes may be generated by throwaway FGD processes.	Has only been applied to Claus process tail gases.
Effects of High CO_2 in Feed	Reduces conversion efficiency of catalyst and decreases H_2S absorption in Stretford solution.	Reduces conversion efficiency of catalyst and efficiency of alkanolamine system.	None.	No effect.
Capital Costs	\$20 to \$60 x 10^3 per Mg/day of S at Claus plant.	\$20 to \$60 x 10^3 per Mg/day of S at Claus plant.	\$40 to \$110 x 10^3 per Mg/day of S at Claus plant.	
General Comments	Exact ppm limit achievable in coal gasification application is not known. Vendor believes 100 ppm is attainable.	Off-gas from amine scrubber is not as low in total sulfur as Beavon process.	On-site boiler/FGD system is the most likely candidate Installing a separate incinerator and FGD would not be as economically feasible.	Much higher residuals in tail gas than Beavon process.

(Continued)

TABLE 4-7. (Continued)

	Cleanair	IFP Claus 1,500	IFP-2	BSR/Selectox
Principle of Operation	Catalytic reduction of sulfur compounds to H_2S followed by a continuation of the Claus reaction and Stretford process.	Liquid phase continuation of Claus reaction at a low temperature.	Incineration of tail gas followed by ammonia scrubbing. Solution is evaporated to produce a concentration SO_2 stream which is returned to the Claus plant.	Catalytic reduction of sulfur compounds to H_2S, followed by oxidation of H_2S to sulfur over Selectox catalyst.
Components Removed	H_2S, COS, CS_2, and SO_2.	H_2S, and SO_2.	COS, CS_2, and H_2S.	H_2S, SO_2, COS, and CS_2.
Efficiency	Reduces sulfur to less than 250 to 300 ppm SO_2 equivalent in effluent.	Reduces sulfur species in Claus tail gas to 1500 ppm as SO_2.	Reduces sulfur species in Claus tail gas to less than 500 ppm.	Up to 99.5% total sulfur removal equivalent to 750 ppmv SO_2 in the incinerated off-gas.
Feed Stream Requirements/ Restrictions	$H_2S:SO_2$ ratio can vary up to 8 to 1 without affecting efficiency; designed specifically for Claus tail gas.	$H_2S:SO_2$ ratio must be maintained in the range of 2.0 to 2.4.	$H_2S:SO_2$ ratio must be maintained in the range of 2.0 to 2.4.	$H_2S:SO_2$ ratio must be maintained in the range of 2.0 to 2.4. HC and NH_3 should not be in the feed.
By-Products	Elemental sulfur.	Elemental sulfur.	Concentrated SO_2.	Elemental liquid sulfur.
Secondary Waste Streams	Spent catalyst.	Spent catalyst.	Spent catalyst.	Spent Beavon and Selectox catalyst, and sour condensate.
Reliability/ Limitations	Has only been applied to Claus process tail gases.	Has only been applied to Claus process tail gases.	Has only been applied to Claus process tail gases.	Has only been applied to Claus plant tail gas.
Effects on High CO_2 in Feed	Reduces conversion efficiency of catalyst; decreases H_2S absorption in Stretford solution.	No effect.	No effect.	Reduces conversion efficiency of BSR catalyst.
General Comments	Cannot attain as low a residual sulfur level in tail gas as Beavon process.	Cannot attain as low a residual sulfur level in tail gas as Beavon process.	Cannot attain as low a residual sulfur level in tail gas as Beavon process.	Higher sulfur emissions than Beavon process.

3) Extensions of the Claus process - such as the IFP-Clauspol 1500 and Sulfreen processes.

Processes in the first category involve catalytic conversion of oxidized sulfur species to H_2S followed by H_2S removal from the gas stream by solvent absorption. In general, the designs of these processes are influenced by high levels of CO_2 in the feed gas. The CO_2 reduces the efficiency of catalytic reduction of COS and CS_2 to H_2S and impairs the effectiveness of the H_2S removal/recovery systems. The SCOT, Beavon, and Wellman-Lord processes are discussed in further detail in Appendices A-8 through A-10.

Both the Beavon and the SCOT processes are commercially available catalytic processes which are potentially applicable to sulfur plant tail gases in EDS commercial plants. These processes feature two sections: a hydrogenation section to convert sulfur species in the gas to H_2S, and an H_2S absorption section. In the hydrogenation reactor of both processes, a reducing gas is added to the feed gas and the combined gas stream is passed over a cobalt molybdate catalyst. The hydrogenation/hydrolysis reactions occur in the catalyst bed converting the sulfur species to H_2S.

The Beavon process employs a Stretford unit for H_2S absorption and elemental sulfur production. In contrast, the SCOT process employs an alkanolamine scrubbing system for H_2S absorption. The absorbing solution in the SCOT is then regenerated, resulting in a H_2S-rich acid gas which is ordinarily returned to the parent Claus plant for treatment. The alkanolamine scrubbing system ultimately limits the SCOT's capabilities because the solvent is only partially selective for H_2S over CO_2. Thus, where feeds to the SCOT contain large amounts of CO_2, it is more difficult to generate an H_2S stream suitable for Claus processing while simultaneously obtaining a SCOT off-gas stream with a low level of total sulfur. For typical applications, vendors of the Beavon process report that levels of less than 100 ppmv total sulfur can be achieved (Appendix A-9), while vendors of the SCOT process report less than 250 ppmv total sulfur (Claus plant tail gas basis) (Appendix A-8). H_2S levels in the Beavon off-gas are often below 10 ppmv; COS is the major contributor to the total sulfur emissions for the Beavon process. By contrast, H_2S levels in the SCOT absorber effluent are typically 200 ppmv. The higher levels of H_2S in SCOT tail gases in existing applications often necessitate that the gases be incinerated in order to minimize odor problems while Beavon tail gases have generally not required incineration. The SCOT process is discussed in detail in Appendix A-8, and the Beavon process in A-9.

The second category of processes involves incineration of the waste gas followed by SO_2 removal. Such processes are capable of achieving levels as low as 150 ppmv of SO_2 in tail gas. One of the prominent SO_2 removal processes, the Wellman-Lord process, removes SO_2 with an alkaline sodium sulfite solution. Subsequent regeneration of the absorbent generates a concentrated SO_2 stream which would be recycled to the parent Claus plant. Eight of the twenty-nine operating Wellman-Lord units have been specially designed to treat Claus plant tail gas (incineration is an integral part of the Wellman-Lord design for this application). However, other FGD processes are also applicable for the removal of SO_2 from incinerated Claus plant tail gas. The Wellman-Lord process for sulfur plant tail gas treatment is discussed in detail in Appendix A-10.

Processes in the third category are used exclusively for Claus plant tail gases, and are capable of no better than 80% recovery of tail gas sulfur. These processes are extensions of the Claus process and therefore require a 2:1 ratio of H_2S to SO_2 for proper operation. The limited sulfur removal capabilities of these processes result in sulfur concentrations of not less than 1000 ppmv. To date, these processes have not been proposed for synthetic fuel plant applications in the U.S.

Incineration

As discussed above,some residual reduced sulfur species will remain in the off-gases leaving many of the bulk sulfur removal and residual sulfur removal processes. Also, these processes achieve only partial control of HCN, CO, NH_3, and organic emissions. The combustion furnace of a Claus unit destroys these constituents, but only in that portion of the gas passing through the furnace. Cyanide can also be destroyed over Claus type catalysts used ahead of the Claus plant itself. HCN is removed from feed gases in the Stretford process, forming SCN^- which leaves the systems with the aqueous blowdown. NH_3 and heavy organics are mostly removed by Stretford solvent, while light organics and CO are not removed by Stretford. Catalytic sulfur tail gas treatment systems achieve at least partial control of any residual HCN contained in sulfur plant tail gases. Both Beavon and SCOT catalytic sections are expected to achieve a high degree of conversion of HCN to NH_3 and CO. Hydrocarbons, CO, and NH_3 contained in the feed to Beavon or SCOT units or added/generated within such units will be present in their tail gases.

Incineration of these tail gases is an effective approach to controlling residual reduced sulfur compounds, hydrocarbons, CO, and NH_3. In addition, incineration is essentially the only alternative for controlling emissions of these pollutants from transient gases such as Streams 801, 802 and 803. SCOT tail gases are ordinarily incinerated to minimize odor problems arising from residual H_2S. Beavon tail gases (with lower H_2S levels) are not ordinarily incinerated, but an incineration step can be added for control of organics, CO, and/or NH_3 if necessary. Sulfur dioxide tail gas treatment processes such as the Wellman-Lord inherently achieve control of organics, CO, and NH_3 as part of the preceding incineration step. Hence, no further control for these constituents is ordinarily necessary with this type of process.

Advantages and disadvantages of the various tail gas and waste gas incineration technologies are summarized in Table 4-8. Generally, a greater degree of control is obtained with high temperature incineration in either a fuel-fired boiler or a separate incinerator (either thermal or catalytic) than can

TABLE 4-8. COMPARISON OF INCINERATION PROCESSES

Type of Incineration	Advantage	Disadvantage	Costs (Total Depreciable Capital)
Thermal Incineration via Separate Incinerator	Can handle all types of waste gases. Reliable and simple operation is common. VOC/CO control and oxidation of sulfur compounds simultaneously.	High supplemental fuel costs for streams with low heating value, control is a problem with streams of varying flow and composition.	On basis of kmols/hr of flow: flow range of 0.3 to 3.0 x 10^3 kmol/hr a. no heat recovery: \$140 to \$870 b. primary heat recovery: \$190 to \$1000 c. primary and secondary heat recovery: \$225 to \$1200
Thermal Incineration in Fuel Fired Boiler	Sulfur and particulates can be removed in the associated electrostatic precipitator and flue gas desulfurization (FGD) units when these are integral with the boiler. The fuel required for steam boiler incineration is less than that of a separate incinerator for wastes with low heating values.	In most cases, this option is more capital intensive than a separate incinerator; however, extent of heat recovery is generally greater with boilers than with incinerators. Subject to control problems with varying waste gas flow rates and compositions.	Incremental boiler capital costs are \$2000-\$3000/kg mole of incremental flue gas compared to coal combustion on a heating value basis. Incremental ESP and FGD costs are an additional \$2000-\$2500/kg mole incremental flue gas.
Catalytic Incineration	Requires less fuel than thermal incineration, although heat recovery may not be as high. Waste gases with very little combustible material can often be incinerated catalytically without supplemental fuel.	Cannot handle large quantities of particulates; they will gradually coat the catalyst and reduce its efficiency. Some catalyst can be easily poisoned by sulfur compounds and elements such as arsenic and lead. High levels of hydrocarbons can raise catalyst to excessive temperatures and shorten the useful life of the catalyst. Temperature control is also a problem with streams of varying flow and composition.	\$400 to \$2200 per kmol/hr of flow for a flow rate range of 40 to 7100 kmol/hr
Flaring	Simple to operate, least expensive alternative, especially for transient and small volume waste gases.	Destruction efficiencies much lower than for thermal or catalytic incineration. Performance data are generally lacking.	• 40 to 100 ft elevated flares for flow rate range of 800 to 6500 kmol/hr - \$25 to \$125 per kmol/hr • Ground flares for flow rate of 100 to 1000 kmols/hr - \$800 to \$2700 per kmol/hr

be achieved through the use of flares, because temperature, residence time, and the amount of air feed are better controlled. The main combustion zone of a gas incinerator is typically engineered to maintain a minimum gas temperature of 1100°K for a minimum of 0.5 seconds. These conditions should result in nearly complete destruction of volatile organic compounds, reduced sulfur compounds, organic aerosols, CO, NH_3, HCN, and particulate matter consisting primarily of combustible material.

Thermal incineration may also be effected using the coal-fired steam or power boiler on site, where a minimum combustion temperature of 1500°K and a minimum residence time of 0.5 seconds are typical design parameters. The coal-fired boiler could be adapted so that the firebox could handle the feed of these tail gas and waste gas streams, and the flue gas system could handle the resulting increase in exhaust gases. This approach results in a degree of pollutant destruction similar to that which would be achieved in a specially engineered incinerator. In tail gas or waste gas treatment applications, using the boiler as an incinerator results in an increase in the boiler capital and annualized operating costs. A larger boiler firebox will be needed, and more coal must be fed for a given steam/power output, due to the requirements for heating the volume of waste gases and the inability to recover all the thermal energy expended. Costs will also increase for the enlarged ESP and FGD units, since costs of the pollution control units are flow rate dependent, and these tail gases and waste gases will not be of sufficient heating value to displace the primary boiler fuel. Detailed discussion of thermal incineration is provided in Appendix A-16.

Catalytic incineration is not likely to be an attractive alternative for control of carbon monoxide and organic emissions in tail gases from sulfur recovery units or transient waste gases due to the presence of sufficient reduced sulfur compounds to interfere with or degrade catalysts. However, catalytic incineration of CO-rich off-gases from acid gas removal (AGR) systems has considerable promise, and is featured in the recent design of at least one U.S. coal gasification facility under construction (based on Texaco gasification). Detailed discussion of catalytic incineration is provided in Appendix A-17.

In the ensuing subsections, the applicability of the control techniques discussed above is addressed for each of the specific waste streams from the EDS base case and MFS case designs covered under Air Source Type 1.

4.2.1.1 Combined Acid Gases (Streams 508, 501, 440)

Details of pollution control alternatives applicable to the following three waste streams in the EDS designs are discussed in this section:

- Stream 508 - acid gas from DEA regenerator
- Stream 501 - acid gas from sour water stripper/ammonia recovery
- Stream 440 - flash gas from partial oxidation unit (MFS case only).

These individual streams are addressed together because, due to their similar characteristics, they will normally be combined in a commercial EDS facility and treated by a single set of controls as a combined acid gas stream. The estimated characteristics for the combined acid gas stream are presented in Tables 4-9 and 4-10 for the base case and MFS case designs, respectively.

The likelihood of combining Streams 508 and 501 for treatment derives from the high H_2S concentrations of both of these streams. Combination of these two acid gas streams for treatment has also been proposed in all known design studies for direct coal liquefaction commercial plants. For the MFS case design, Stream 440 has been included in the combined acid gas stream due to the relatively high COS concentration (1180 ppmv) of this stream. The high COS concentration in Stream 440 means that it cannot be conveniently treated in conventional Stretford units, which do not remove any of the COS present.

The control approach for the combined acid gas stream involves all three functions described previously: bulk sulfur removal, residual sulfur removal, and incineration.

TABLE 4-9. ESTIMATED CHARACTERISTICS OF COMBINED ACID GASES FOR EDS COMMERCIAL PLANT (ILLINOIS COAL BASE CASE)*

Component	Stream 508, kmol/hr	Stream 501, kmol/hr	Combined Acid Gases	
			Volume %	kmol/hr
CO_2	312.1	65.4	25.56	377.5
H_2S	779.4	142.3	62.40	921.7
NH_3	47.7	1.8[†]	3.35	49.5
COS	0.45	0	0.03	0.45
H_2O	74.2	53.8	8.66	128.0
Total	1,213.9	263.3	100.00	1,477.2
Temperature, °K	322	322		322
Pressure, MPa	0.2	0.1		0.1

* EDS plant producing 9,580 m^3 (60,240 barrels) fuel oil equivalent of liquid products per stream day.

[†] Assumes NH_3 recovery is employed in the sour water stripping operation.

Note: The single composition numbers used here (rather than ranges) reflect a single set of design and operation conditions for the EDS plant. In practice, compositions could vary due to uncertainty in these estimates, and due to variations in coal type, design parameters, and operating conditions. This note is equally applicable to other composition tables or any single composition numbers presented in Section 4.

TABLE 4-10. ESTIMATED CHARACTERISTICS OF COMBINED AND ACID GASES FOR EDS COMMERCIAL PLANT (ILLINOIS COAL MFS CASE)*

Component	Stream 508, kmol/hr	Stream 501, kmol/hr	Stream 440, kmol/hr	Combined Acid Gases	
				Volume %	kmol/hr
CO_2	296.8	55.1	82.7	25.53	434.6
H_2S	772.5	136.8	11.2	54.07	920.5
NH_3	62.2	1.6[†]	0.04	3.75	63.8
COS	0.45	0	0.3	0.044	0.75
H_2O	73.7	49.7	19.0	8.36	142.4
H_2	0	0	61.3	3.60	61.3
C_1	0	0	0.08	0.0047	0.08
N_2	0	0	1.4	0.082	1.4
CO	0	0	77.7	4.56	77.7
Total	1,205.7	243.2	253.7	100.00	1,702.6
Temperature, °K	322	322	322		322
Pressure, MPa	0.2	0.1	0.2		0.1

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* EDS plant producing 11,300 m^3 (71,080 barrels) fuel oil equivalent of liquid and gaseous products per stream day.

[†] Assumes NH_3 recovery is employed in the sour water stripping operation.

4.2.1.1.1 Control Function 1 - Bulk Sulfur Removal

In an EDS commercial plant, the combined acid gas streams contain 62.4% H_2S and 54.1% H_2S for the base case and MFS case design, respectively. Such a stream is highly suitable for bulk sulfur removal by the Claus process, from both technical and economical considerations, because it has such a relatively high H_2S concentration. Other available bulk sulfur removal processes generally operate best on streams with lower H_2S levels, and hence would be less appropriate for the treatment of this gas stream.

4.2.1.1.1.1 Control Technique 1 - Claus Process

Because of the high H_2S concentration in the combined acid gas stream, Claus plant designs of the "straight-through" mode are a reasonable choice for bulk sulfur removal. The "straight-through" mode of operation has several distinct advantages. Most of the organics and CO present would be combusted, because organics and CO tend to be preferentially oxidized before one-third of the H_2S is oxidized to SO_2. Likewise, the majority of the ammonia present would be preferentially burnt to nitrogen and water, and some NO_x. The presence of ammonia is of some concern because the levels of ammonia in the combined acid gas stream, at 3.4% for the base case and 3.8% for the MFS case, are considered to be marginally acceptable as Claus plant feed. If ammonia is not properly combusted, residual ammonia would react with hydrogen sulfide, sulfur dioxide, and carbon dioxide to form salts (e.g., ammonium hydrosulfide, ammonium polysulfides, ammonium carbonates or carbamates and/or ammonium sulfate) that precipitate throughout the Claus unit, causing plugging and catalyst deactivation problems which frequently result in premature plant shutdowns (Appendix A-6). Thus, special attention in the design of burners and combustion chambers and proper control of air feed are necessary for the successful operation of the Claus unit in the EDS commercial plant. If ammonia is not recovered after stripping of the sour water, Stream 501 will contain high levels of NH_3 and a two compartment combustion chamber might be employed for the Claus design. The NH_3-bearing stream would be fed to the first compartment along with a sub-stoichiometric amount of air. Operation under these reducing conditions would lower the tendency to form SO_3, while hardly diminishing the destruction of NH_3. The other acid gas streams would then be fed to the second compartment of the combustion chamber, along with additional air, and allowed to react with the SO_2 formed in the first compartment to produce sulfur.

As discussed previously, Claus units of 3-stage design can achieve overall sulfur removal efficiencies of over 95%. Assuming a 95% sulfur removal efficiency, a Claus tail gas sulfur loading of 46.6 kmol/hr and 46.5 kmol/hr would be expected for the base case and MFS case designs, respectively.* The gaseous sulfur species distribution in Claus tail gas is approximately 60% H_2S, 30% SO_2, 9% COS and 1% CS_2. Up to 50% of the Claus tail gas total sulfur (SO_2 equivalent basis) may actually be present as either sulfur vapor or entrained elemental sulfur. However, this would not appreciably alter the tail gas sulfur species distribution (e.g., for 50% SO_2 equivalent sulfur as S_8, the species distribution would become 53% H_2S, 27% SO_2, 8% COS, 1% CS_2, and 11% S_8). The corresponding concentration of total sulfur (SO_2 equivalent) in the Claus tail gas would be 13,600 ppmv and 11,900 ppmv for the base case and MFS case designs, respectively. The corresponding total tail gas flow rate would be 3,420 kmol/hr and 3,930 kmol/hr.

Unit cost and capital investment costs for Claus plants were obtained from the cost curve provided in Appendix A-6; total process cost (TPC,sum of installed equipment cost and indirect installation cost) for the Claus plant (2 parallel Claus trains; 14.8 Mg/hr or 461 kmol/hr sulfur per Claus train for either the base case or the MFS case) was obtained by reading from Figure A 6-5 in Appendix A-6 and multiplying by a factor of two (for 2 parallel trains); this cost was converted to total capital investment (TCI) using the procedures shown in Table 4-1. Annualized costs for the Claus process were derived using the methodology described in Section 4.1.3 (summarized in Table 4-2), and based on utilities and operating labor requirements provided in Appendix A-6. For a Claus plant in the capacity range designed for an EDS commercial plant, the capital investment cost per unit capacity of bulk sulfur removal would be about $0.97 million per Mg/hr sulfur recovered. For the base case design with no recycle sulfur stream from tail gas treatment, the total capital investment cost would be $27.1 million (assuming 2 parallel trains) and the total annualized cost would be $3.0 million.* These costs would be the same for the MFS case design. The annualized cost is low due to credits for steam (about $2.0 million) generated by the exothermic Claus reactions.

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* Material flows are also presented in tabular form in Section 4.2.1.6 (Tables 4-18 and 4-19) under Integrated Control Examples.

Secondary Waste Streams

The only secondary waste stream generated by the Claus process is spent catalyst. The EDS Claus plant would have a bauxite or alumina catalyst inventory of approximately 206 Mg and 237 Mg for the base case and MFS case designs, respectively. Assuming a catalyst life of 5 years, the average spent catalyst generation rate would be 41 Mg per year and 47 Mg per year for these two designs. The spent catalyst consists of bauxite or alumina contaminated by the buildup of carbonaceous matter, elemental sulfur, and sulfates. Controls for this secondary waste stream are addressed in Section 4.4.5.

Depending on the tail gas treatment process employed, the amount of sulfur recovered in the Claus plant would range from 28.1 to 29.5 Mg/hr. The higher sulfur recovery corresponds to tail gas treatment processes (such as SCOT) with sulfur streams recycled to the Claus plant.

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* Costs are also presented in tabular form in Section 4.2.1.6 (Table 4-20) under Integrated Control Examples.

4.2.1.1.2 Control Function 2 - Residual Sulfur Removal

In an EDS commercial plant, tail gas from bulk sulfur removal would contain 13,600 ppmv and 11,900 ppmv total sulfur for the base case and MFS case designs, respectively. Traces of NH_3 may also be present as a result of incomplete NH_3 destruction. A number of processes are available for recovering residual sulfur from the bulk sulfur removal tail gas (refer to Section 4.2.1). For purposes of discussion, the Beavon, SCOT and Wellman-Lord processes will be considered.

4.2.1.1.2.1 Control Technique 1 - Beavon Process

The Beavon process is reported to be capable of reducing residual sulfur concentrations to about 100 ppmv with less than 10 ppmv H_2S (Appendix A-9). Available data show that the Beavon off-gas often contains significantly less than 100 ppmv of total sulfur compounds. Data from four Beavon units tested indicate that the level of H_2S in the off-gas ranged from 0.1 - 7 ppmv, whereas the levels of COS and CS_2 ranged from 5 - 50 ppmv and from 0.1 - 0.2 ppmv, respectively (Appendix A-9). Assuming a 100 ppmv sulfur level in the Beavon off-gas, the total sulfur emission rate would be 0.30 kmol/hr and 0.35 kmol/hr for the base case and MFS case designs, respectively. The corresponding amount of sulfur recovered in the Beavon unit would be 1.46 Mg/hr for both EDS designs.* An external source of reducing gas may be required for the catalytic hydrogenation of SO_2 and other sulfur species prior to Stretford absorption. Reducing gas may be added directly to tail gas treatment in the form of H_2 and CO-rich gas or may be generated by substoichiometric combustion of fuel gas derived from synthesis operations. The total process cost (TPC) for the Beavon plant (2 parallel Beavon trains; each train removes 17.5 Mg/day sulfur for the base case design) was obtained by reading from Figure A 9-2 in Appendix A-9 and multiplying by a factor of two (for 2 parallel trains); this cost was converted to total capital investment (TCI) using the procedures shown in Table 4-1. Annualized costs for the Beavon process were derived using the methodology described in Section 4.1.3, utilities and chemicals requirements provided in Appendix A-9, and an operating labor requirement of 0.25 man/shift. For the base case design, the total capital investment cost (assuming 2 parallel trains) and the total annualized cost would be $21.1 million and $8.5 million, respectively.* These costs would be approximately the same for the MFS case design. For a Beavon plant in the capacity range designed for an EDS commercial plant, the unit capital investment cost (TCI) would be approximately $14 million per Mg/hr sulfur removed in tail gas treatment.

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* Materials flows and costs are also presented in tabular form in Section 4.2.1.6 (Tables 4-18, 4-19 and 4-20) under Integrated Control Examples.

Secondary Waste Streams

Secondary waste streams from the Beavon process are: 1) sour reactor effluent condensate; 2) Stretford solution purge; 3) Stretford oxidizer vent gas; and 4) spent catalyst from the Beavon hydrolysis reactor. Sour condensate is generated by the condensation of water vapor present in the Beavon hydrolysis reactor effluent. This sour condensate is expected to contain dissolved sulfide and traces of NH_3, and would be generated at a rate of 17.9 m^3/hr and 19.4 m^3/hr for the base case and MFS case designs, respectively. Controls for this secondary waste stream are identical to those for sour water streams addressed in Section 4.3.1. Stretford solution purge for control of thiosulfate buildup typically contains 0.19 wt % anthraquinone disulfonic acid (ADA), 0.38 wt % vanadium, 1.45 wt % Na_2CO_3, 8.95 wt % $Na_2S_2O_3$, and 4.02 wt % Na_2SO_4 (Appendix A-9). The solution purge rate is approximately 441 kg/hr for both EDS designs. As will be discussed in Section 4.2.1.2, and also in Appendix A-7, this purge stream may be regenerated by reductive incineration or discarded. Oxidizer vent gas would consist primarily of air, water vapor, and CO_2, but may contain traces of NH_3 and probably some entrainment of the Stretford solution. Controls for this secondary waste stream, if required, are addressed in Section 4.2.1.2.1.1. Insufficient data are available for estimating the flow rate of oxidizer vent gas. The cobalt molybdate hydrogenation catalyst inventory is approximately 59 Mg for both EDS designs and would require periodic replacement. Assuming a catalyst service life of three years, the average spent catalyst generation rate would be 20 Mg/year. The spent catalyst consists of cobalt and molybdenum contaminated with sulfur and sulfur compounds. Controls for the spent catalyst are addressed in Section 4.4.5.

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* Materials flows and costs are also presented in tabular form in Section 4.2.1.6 (Tables 4-18, 4-19 and 4-20) under Integrated Control Examples.

4.2.1.1.2.2 Control Technique 2 - SCOT Process

The SCOT process is reported to be capable of reducing residual sulfur concentrations to about 250 ppmv with approximately 200 ppmv of H_2S (Appendix A-8). This corresponds to a total sulfur emission rate of 0.79 kmol/hr and 0.93 kmol/hr for the base case and MFS case designs, respectively.* Sulfur is not recovered in the SCOT process; a concentrated H_2S stream is recycled to the Claus plant for sulfur recovery. The total process cost (TPC) for the SCOT plant (2 parallel SCOT trains; each train treats off-gas from a 373 Mg/day sulfur Claus train and removes 0.77 Mg/hr sulfur for the base case design) was obtained by reading from the upper cost curve in Figure A 8-2 of Appendix A-8 and multiplying by a factor of two (for 2 parallel trains); this cost was converted to total capital investment (TCI) using the procedures shown in Table

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* Material flows and costs are also presented in tabular form in Section 4.2.1.6 (Tables 4-21, 4-22 and 4-23) under Integrated Control Examples.

4-1. Annualized costs for the SCOT process were derived using the methodology described in Section 4.1.3, and utilities, chemicals and operating labor requirements provided in Appendix A-8. For the base case design, the total capital investment cost (assuming 2 parallel trains) and the total annualized cost would be \$18.4 million and \$7.9 million, respectively.* These costs would be approximately the same for the MFS case design. For a SCOT plant in the capacity range designed for an EDS commercial plant, the unit capital investment cost (TCI) would be approximately \$12 million per Mg/hr of sulfur removed in tail gas treatment.

Secondary Waste Streams

Secondary waste streams from the SCOT process are sour water and spent catalyst. The sour water is generated from condensation of the water vapor present in the SCOT hydrolysis reactor. This sour water is expected to contain dissolved sulfide and traces of NH_3, and would be generated at a rate of 18.8 m^3/hr and 20.3 m^3/hr for the base case and MFS case designs, respectively. Controls for this sour water stream are identical to those for sour water streams addressed in Section 4.3.1. The cobalt molybdate hydrogenation catalyst inventory is approximately 59 Mg for both EDS designs. Assuming a catalyst service life of 5 years, the average spent catalyst generation rate would be 12 Mg/yr. The spent catalyst consists of cobalt and molybdenum contaminated with sulfur and sulfur compounds. Controls for the spent catalyst are addressed in Section 4.4.5.

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* Material flows and costs are also presented in tabular form in Section 4.2.1.6 (Tables 4-21, 4-22 and 4-23) under Integrated Control Examples

4.2.1.1.2.3 Control Technique 3 - Wellman-Lord Process*

The Wellman-Lord SO_2 recovery process is typically designed to achieve effluent levels of less than 150 ppmv to 250 ppmv SO_2. Effluent levels of less than 100 ppmv SO_2 in the stack gas have been consistently achieved in commercial installations (Appendix A-10). More recently, Davy McKee has designed a Wellman-Lord SO_2 absorption system to treat the Claus tail gas in a coal liquefaction plant, so as to reduce the SO_2 concentration in the treated stack gas to a maximum of 150 ppmv (Appendix A-10). Unlike the Beavon and SCOT process, residual sulfur compounds in the Claus tail gas are incinerated to SO_2 prior to absorption in the Wellman-Lord process. Thus, the composition and heating value of the fuel gas used for Claus tail gas incineration would impact the total flow rate of the gas stream entering the Wellman-Lord absorption system. Use of low-Btu fuel gas would result in higher flow rate of combustion product gas entering the Wellman-Lord absorber. And, at the higher flow rates, the 150 to 250 ppmv SO_2 level achievable with the Wellman-Lord process translates into higher SO_2 mass emissions.

For the EDS base case and MFS case designs, an effluent concentration of 150 ppmv SO_2 corresponds to a total sulfur emission rate of 1.40 kmol/hr and 1.63 kmol/hr, respectively.† These emission rates were estimated with the assumption that the low-Btu fuel gas from Flexicoking would be used for incineration of the Claus tail gas. Sulfur is not recovered in the Wellman-Lord process; a concentrated SO_2 stream is recycled to the Claus plant for sulfur recovery. The capital investment cost per unit capacity of the Wellman-Lord unit would be approximately $22 million per Mg/hr of sulfur removed in tail gas treatment. For the base case design, the total capital investment cost (assuming two parallel incinerator-absorber trains and one regenerator) and the total annualized cost would be \$33.8 million and \$11.1 million, respectively.* Again, these costs would be approximately the same for the MFS case design. The capital investment costs for the Wellman-Lord process were derived from the sum of the total process cost (TPC) curves for the incinerator-absorber train (2 parallel trains) and the regenerator (1 train) provided in Appendix A-10. Incineration is an integral part of the Wellman-Lord process and included in the cost estimate. Annualized costs for the Wellman-Lord process were derived using the methodology described in Section 4.1.3, and utilities, chemicals and operating labor requirements provided in Appendix A-10.

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† Material flows and costs are also presented in tabular form in Section 4.2.1.6 (Tables 4-24, 4-25 and 4-26) under Integrated Control Examples.

* Note that incineration of bulk sulfur removal tail gases, prior to W-L absorption, is an integral part of the Wellman-Lord process.

Secondary Waste Streams

Secondary waste streams from the Wellman-Lord process are acidic wastewater from combustion gas quenching and thiosulfate/sulfate by-product purge. The acidic wastewater is generated from condensation of water vapor present in the incinerator off-gas, and typically has a pH value between 1 and 2 due to the SO_2 absorbed. No other composition data for the acidic wastewater are available. This wastewater would be generated at a rate of 67.6 m^3/hr and 76.5 m^3/hr for the base case and MFS case designs, respectively. Controls for this wastewater stream are identical to those for Water Source Type 1 addressed in Section 4.3.1. The by-product purge typically contains 41.5 wt % Na_2SO_3, 18.1 wt % $Na_2S_2O_5$, 10.0 wt % Na_2SO_4, 1.1 wt % $Na_2S_2O_3$, and 29.2 wt % H_2O (Appendix A-10). By-product purge is generated at a rate of 314 kg/hr for both EDS designs. Controls for this purge stream are identical to those for Solid Waste Source Type 1 addressed in Section 4.4.2.

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* Material flows and costs are also presented in tabular form in Section 4.2.1.6 (Tables 4-24, 4-25 and 4-26) under Integrated Control Examples.

4.2.1.1.3 Control Function 3 - Incineration

Waste gases from tail gas treatment may be incinerated to control odors due to the presence of H_2S and to remove any residual organics and CO. As discussed in Section 4.2.1, thermal incineration, catalytic incineration and flaring are employed in various commercial applications. Thermal incineration in a dedicated incinerator is an approach often used in conjunction with tail gas treatment units, and will be considered in this section.

4.2.1.1.3.1 Control Technique 1 - Thermal Incineration

In an EDS commercial plant, gases from tail gas treatment processes (such as Beavon and SCOT) not having an integral incineration step are generated at a rate of approximately 3,150 kmol/hr and 3,710 kmol/hr for the base case and MFS case designs, respectively. The sulfur content of these gases might range from 100 to 250 ppmv with H_2S concentrations ranging from less than 10 to 200 ppmv. The level of combustibles is expected to be inadequate to support combustion, and supplemental fuel would be required for incineration. Using treated low-Btu fuel gas from Flexicoking and assuming an incineration temperature of 978°K (1300°F) and 20% excess air for combustion, about 6,570 kmol/hr and 7,730 kmol/hr of incinerated gas (including combustion products from low-Btu fuel gas) would be generated for the base case and MFS case designs, respectively. The concentration of SO_2 in the effluent would be approximately 140 ppmv. Concentrations of CO and hydrocarbons in the incinerated gas would typically range from 10 - 100 ppmv and 5 - 20 ppmv (Appendix A-16). No secondary waste streams are generated by thermal incineration processes. The capital investment cost per unit capacity of the thermal incinerator with waste heat recovery would be approximately \$940 - \$1,010 per kmol/hr of gas from tail gas treatment, for gas flows in the 3,150 - 3,710 kmol/hr range. For the base case design, the total capital investment cost and the total annualized cost for SCOT off-gas incineration with waste heat recovery would be \$3.2 million and \$4.7 million, respectively. The corresponding costs for the MFS case design would be \$3.5 million and \$5.5 million. The capital investment costs for thermal incineration were obtained using the cost curve provided in Appendix A-16. Annualized costs for thermal incineration were derived using the methodology described in Section 4.1.3, and based on calculated fuel requirements and steam generation rates from energy balances and an operating labor requirement of 0.125 man per shift.

4.2.1.2 Acid Gas from Acid Gas Removal Unit in Hydrogen Purification (Stream 428)

This gaseous waste stream results from regeneration of the solution from the Catacarb process (or other acid gas removal systems) used in removing H_2S and CO_2 from the shift gas following partial oxidation, and applies only to the MFS case design. The stream is generated at a rate of 13,304 kmol/hr, and contains 1.03% H_2S, 38 ppmv COS, 669 ppmv NH_3, 23 ppmv C_1, 60 ppmv CO, and 91.18% CO_2. The low sulfur level and high flow rate imply that this stream might not be suitably combined with the other acid gas streams (Streams 508, 501, and 440) for treatment. Further, the low levels of hydrocarbons and CO present indicate that incineration to control these pollutants might not be so critical. For this particular waste stream, the only control function needed appears to be removal of the H_2S present by bulk sulfur removal. Because of the high levels of CO_2 present, however, treatment of this gas stream by residual sulfur removal processes such as Beavon and SCOT would not be effective. The high CO_2 levels would result in high equilibrium COS levels in the Beavon or SCOT hydrolysis reactions. The COS formed would not be removed by the Stretford absorber in the Beavon process; thus, the addition of the hydrolysis step would result in an effluent containing higher sulfur levels than the effluent from the Stretford process alone (without the hydrolysis step). For the SCOT process, a portion of the CO_2 would be co-absorbed with the H_2S and COS. Regeneration of the amine absorption solution would result in an H_2S-lean acid gas stream consisting primarily of CO_2, and unsuitable for further processing in Claus type sulfur recovery units. Incineration of Stream 428 followed by SO_2 removal is an alternative that might be considered; however, this alternative is not economically attractive for application to this particular waste stream when compared with bulk sulfur removal processes such as the Stretford, because of the low levels of combustibles in Stream 428 and the supplemental fuel requirement for incineration of this gas stream.

4.2.1.2.1 Control Function 1 - Bulk Sulfur Removal

The low H_2S level in the acid gas stream suggests that the Stretford or the Giammarco-Vetrocoke process would be suited for controlling sulfur emissions from this stream. For purposes of discussion, only the Stretford process will be considered.

4.2.1.2.1.1 Control Technique 1 - Stretford Process

The Stretford process has generally been proposed for the treatment of sour gas streams containing only a few percent H_2S in coal liquefaction facilities. Concentrations of H_2S in the purified gas stream can typically be reduced to less than 10 ppmv, and sometimes to less than 1 ppmv. Most of the Stretford installations, however, have been designed to treat gases containing only a few percent CO_2. With a feed gas containing over 90% CO_2 as in this case, there is some concern that CO_2 absorption by the Stretford solution would lower the pH and increase the bicarbonate/carbonate ratio, which in turn would lower the rate of H_2S absorption. At the SASOL I Stretford plant, original design and operational problems in treating the high CO_2 (>90%) feed gas had resulted in H_2S levels of 0.8% in the inlet stream and 0.81% in the effluent stream, indicating no H_2S removal. According to Peabody-Holmes, one of the licensees for the Stretford process, these problems could be solved by considering the different rates of CO_2 and H_2S absorption in the Stretford solution. The Peabody-Holmes Stretford design subsequently installed at SASOL I incorporated the following changes: 1) contact of gas and liquid in a venturi scrubber to minimize contact time and hence CO_2 absorption; and 2) modification of the Stretford solution. Peabody-Holmes, for example, reportedly has a Stretford unit operating successfully in St. Elmo, Illinois since 1973 on a stream consisting of 85% CO_2 and 15% H_2S (Appendix A-7). Thus, treatment of sour gas streams containing over 90% CO_2 might warrant special design considerations but does not appear to be a formidable problem.

Air Source Type 1
Acid Gas from H_2
Purification
Control Function 1 -
Bulk Sulfur Removal

The Stretford process is not capable of removing carbonyl sulfide. Assuming a 10 ppmv H_2S level in the Stretford absorber effluent, the total sulfur emitted would include 0.13 kmol/hr of H_2S and 0.50 kmol/hr of COS (representing the 38 ppmv COS in the feed to the Stretford unit). Sulfur recovered from the Stretford plant amounts to 4.4 Mg/hr. The total capital investment cost and the total annualized cost for the Stretford plant are presented in Table 4-11. Both costs represent approximately 1% of the corresponding uncontrolled base plant costs.

Secondary Waste Streams

There are two secondary waste streams from the Stretford plant: 1) Stretford solution purge stream due to thiosulfate and sulfate formation; and 2) oxidizer vent gas. The solution purge is generated at a rate of 1,310 kg/hr, and would typically contain 0.19 wt % anthraquinone disulfonic acid (ADA), 0.38 wt % vanadium, 1.45 wt % Na_2CO_3, 8.95 wt % $Na_2S_2O_3$, 4.02 wt % Na_2SO_4, and 85.01 wt % H_2O. In the past, disposal of this solution purge was often considered. In 1973, a reductive-incineration process was developed which converts the solution purge into a gas stream containing H_2S, water vapor and a solid residue containing soda ash and reduced vanadium salts. The salts are returned to the Stretford process as make-up chemicals and the H_2S-rich gas and water vapor are recycled to the absorber. Thus, the reductive incineration process recovers expensive chemicals while effectively attaining a "zero" discharge of solution purge. The reductive-incineration process is considered to be an integral part of the Stretford process offered by Peabody-Holmes.

The Stretford oxidizer vent gas would contain mostly nitrogen, oxygen, essentially all the ammonia present in the feed gas, and be saturated with water. Assuming that the efficiency for transfer of oxygen from air to the Stretford solution for regeneration is only 1/3, the oxidizer vent gas would be generated at a rate of approximately 1,000 kmol/hr. At this flow rate, ammonia contained in the oxidizer vent gas would be 8,900 ppmv, if all the

TABLE 4-11. COSTS OF BULK SULFUR REMOVAL WITH STRETFORD PROCESS FOR ACID GAS FROM AGR UNIT IN HYDROGEN PURIFICATION (STREAM 428)

Cost Element	Total Capital Investment*	Total Annualized Cost†
Total Cost	$28.1 million	$8.5 million
% of Base Plant‡	0.97	0.94
Unit Cost	$ 6.4 million per Mg/hr sulfur recovered	$245/Mg sulfur recovered

* The total process cost (TPC) was calculated using: 1) 2 parallel Stretford trains, each processing 52.7 Mg/day sulfur; 2) a TPC of $7.8 million for a Stretford unit processing 27.7 Mg/day sulfur (115); 3) cost increases that vary with the 0.6 power of the capacity ratio. The TPC was converted to total capital investment cost (TCI) using the procedures outlined in Table 4-1.

† The annualized cost was derived using methodology described in Table 4-2, utilities and chemicals requirements provided in Table A 7-3 of Appendix A-7 (assuming these requirements are linearly proportional to the amount of sulfur processed), and an operating labor requirement of 0.5 man per shift.

‡ Cost estimates for uncontrolled base plant for both the EDS base case and MFS case designs are presented in Section 2.3.

Note: The single numbers (as versus ranges) presented in this table represent a specific set of design and operating conditions for the EDS plant and for the control technique. Both cost and performance figures could vary due to uncertainties in cost and performance estimates and variations in such parameters as coal type, design and operating conditions. This note is equally applicable to other stream-specific tables in Section 4.

ammonia in the feed to the Stretford appears in this vent gas. The ammonia in the oxidizer vent gas can be readily removed by water scrubbing. Because of the high solubility of ammonia in water, only a few stages in a tray tower, a short packed-bed tower, or a small pressure drop in a venturi scrubber would be needed for ammonia removal. The added cost for incorporating ammonia removal would be a very small fraction of the cost for the Stretford plant presented in Table 4-11. For example, the capital investment cost for a packed-bed tower to remove ammonia from the oxidizer vent gas would be approximately \$70,000. This cost was calculated assuming a packed-bed depth of 1.8 m (adequate for 99 percent removal of ammonia) and using cost information provided by Hanf and MacDonald (44). The wastewater stream from ammonia scrubbing might be sent to the sour water stripper/ammonia recovery system for treatment. Alternately, the oxidizer vent gas might also be used as air feed to boilers, process heaters, incinerators, or the Claus plant, where most of the ammonia would be combusted to form nitrogen and water.

4.2.1.3 Flexicoking Gasifier/Heater Sour Fuel Gas (Stream 304)

The sour fuel gas is generated in the Flexicoking gasifier vessel where coke formed in the Flexicoking reactor is gasified with steam and air. Total sour fuel gas produced is 54,781 kmol/hr and 27,224 kmol/hr for the base case and MFS case designs, respectively. The sour fuel gas from base case design contains 0.42% H_2S, 108 ppmv COS, 11.6% CO_2, 17.9% H_2, 18.0% CO, 1.5% C_1, 47.0% N_2, and 3.6% H_2O. The sour fuel gas from the MFS case design is slightly different in composition in that it contains 92 ppmv COS, 19.7% H_2, and 45.8% N_2. With a heating value of 4.7 MJ/Nm^3 (125 Btu/SCF, dry basis), combustion of this sour fuel gas would result in SO_2 emissions of approximately 2,550 ng/J (5.9 lb/MM Btu). Hence, removal of reduced sulfur compounds from the sour fuel gas prior to combustion in process heaters and boilers is an option that might be considered. The other control option, removal of SO_2 after combustion, is more conveniently discussed under control technologies for combustion gases (Source Type 2).

Two general types of control strategies to remove reduced sulfur compounds from the sour fuel gas are available. The first type involves two control functions: acid gas removal (AGR), and bulk sulfur removal of the acid gas stream from the regenerator of the AGR unit. For example, an AGR system that have a greater affinity for H_2S than for CO_2, such as the Selexol (Appendix A-3) or the Rectisol (Appendix A-2) process, can be used to remove the H_2S along with some of the COS and CO_2 (e.g., 15% of the CO_2) from the sour fuel gas. The acid gas stream from the regenerator will contain practically all of the H_2S, and could be sufficiently concentrated in H_2S -- depending on the AGR process and the design conditions -- to be fed directly to the Claus plant for bulk sulfur recovery. In cases where lower H_2S concentrations are obtained in the acid gas stream from the AGR regenerator, this acid gas stream might be combined with acid gases from other process areas prior to bulk sulfur removal in a Claus plant. The second type of control strategy is to remove and recover reduced sulfur compounds from the sour fuel gas within a single control function. This second type of control strategy is discussed in the following section.

4.2.1.3.1 Control Function 1 - Bulk Sulfur Removal

Since the sour fuel gas only contains approximately 0.4% H_2S, the Stretford or Giammarco-Vetrocoke process would be the choice for reducing the sulfur content of this stream. For purposes of discussion, only the Stretford process will be considered.

4.2.1.3.1.1 Control Technique 1 - Stretford Process

As discussed previously, concentrations of H_2S in the treated fuel gas stream from the Stretford process can generally be reduced to less than 10 ppmv, and sometimes to less than 1 ppmv. The Stretford process, however, is not capable of removing carbonyl sulfide. Thus, the treated fuel gas would be expected to contain less than 10 ppmv H_2S and 108 ppmv COS for the base case design, and less than 10 ppmv H_2S and 92 ppmv COS for the MFS case design. Because of the relatively low CO_2 levels in the sour fuel gas, no specific operational problems involving CO_2 absorption and pH lowering are expected. Sulfur recovered from the Stretford plant would amount to 7.3 Mg/hr and 3.9 Mg/hr for the base case and MFS case designs, respectively.

The total capital investment costs and the total annualized costs for the Stretford plant are presented in Table 4-12. Because three Flexicoking trains have been proposed in the EDS base case design and only two in the MFS case design, the capital investment costs were estimated based on three parallel Stretford units for the base case design and two parallel Stretford units for the MFS case design (i.e., one Stretford unit for each Flexicoker). Both the capital investment cost and the annualized cost represent approximately 1.4% of the corresponding uncontrolled base plant costs for the base case design, and approximately 0.9% of the corresponding uncontrolled base plant costs for the MFS case design.

TABLE 4-12. COSTS OF BULK SULFUR REMOVAL WITH STRETFORD PROCESS FOR FLEXICOKING HEATER/GASIFIER SOUR FUEL GAS (STREAM 304)

Cost Element	Base Case		MFS Case	
	Total Capital Investment*	Total Annualized Cost †	Total Capital Investment*	Total Annualized Cost†
Total Cost	$45.0 million	$13.8 million	$26.1 million	$7.8 million
% of Base Plant	1.41	1.43	0.90	0.86
Unit Cost	$ 6.1 million per Mg/hr sulfur recovered	$237/Mg sulfur recovered	$ 6.7 million per Mg/hr sulfur recovered	$254/Mg sulfur recovered

* The total process cost (TPC) was calculated using: 1) a TPC of $7.8 million for a Stretford unit processing 277 Mg/day sulfur (115); and 2) cost incrases that vary with the 0.6 power of the capacity ratio. The TPC was converted to total capital investment cost (TCI) using the procedures outlined in Table 4-1. For the base case, the capital investment cost is the sum of the costs of three Stretford units, each treating 1/3 of Stream 304 or 58.8 Mg/day sulfur. For the MFS case, the capital investment cost is the sum of the costs of two Stretford units, each treating 1/2 of Stream 304 or 46.6 Mg/day sulfur.

† The annualized costs were derived using the methodology described in Table 4-2, utilities and chemicals requirements provided in Table A 7-3 of Appendix A-7 (assuming these requirements are linearly proportional to the amount of sulfur processed), and an operating labor requirement of 0.5 man per shift.

Secondary Waste Streams

As in other Stretford designs, there are two secondary waste streams from the process: 1) Stretford solution purge stream due to thiosulfate and sulfate formation; and 2) oxidizer vent gas. The solution purge is generated at a rate of 2,186 kg/hr and 1,154 kg/hr for the base case and MFS case designs, respectively. The composition of this purge stream has been described in Section 4.2.1.2. The solution purge might be treated by reductive-incineration with all the sulfur released as H_2S and recycled to the Stretford absorber.

The Stretford oxidizer vent is not expected to cause any problems. This is because of the absence of ammonia and mercaptans in the sour fuel gas. The Stretford oxidizer vent would therefore contain only nitrogen, oxygen, and be saturated with water.

4.2.1.4 Slurry Drier Vent Gas (Stream 102) and Vacuum Fractionator Off-Gases (Streams 153/156)*

The slurry drier vent gas is generated at a rate of 532 kmol/hr, and contains 30 ppmv H_2S and 4.1% hydrocarbons. The vacuum fractionator off-gas is generated at a rate of 69 kmol/hr, and contains 5.0% H_2S and 47.5% hydrocarbons. The slurry drier vent gas has a heating value of approximately 7.1 MJ/Nm^3 (190 Btu/SCF), and might be used as fuel gas without additional treatment. The vacuum fractionator off-gas has a higher heating value but also higher H_2S level. In practice, it might be convenient to combine these two gas streams with the Flexicoking sour fuel gas, so that removal of reduced sulfur compounds from this combined fuel gas stream prior to combustion might be considered. The slurry drier vent gas and vacuum fractionator off-gas together are generated at a rate of 601 kmol/hr, and contain 0.58% H_2S and 9.1% hydrocarbons. Thus, the H_2S level of these two streams together is similar to the H_2S level in the Flexicoking sour fuel gas. The only control function needed for these two streams is bulk sulfur removal. The other control option, removal of SO_2 after combustion, is more conveniently discussed under control technologies for combustion gases (Source Type 2).

Similar to the Flexicoking sour fuel gas (Section 4.2.13), two types of control strategies to remove reduced sulfur compounds from Streams 102 and 153/156 are available. The first type involves acid gas removal (AGR) followed by bulk sulfur removal. AGR techniques that have a greater affinity for H_2S than CO_2 would be applicable if these streams were combined with the Flexicoking sour fuel gas. If Streams 102 and 153/156 were not combined with the Flexicoking sour fuel gas, CO_2 is not present in these streams and non-selective AGR systems such as the DEA process might be considered for sulfur removal.

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* Streams 153/156 in the MFS case design is identical to Stream 153 in the base case design.

The application of AGR techniques to these streams is not discussed here; descriptions of AGR systems can be found in Appendices A-1 through A-5. The second type of control strategy involves removing and recovering reduced sulfur compounds within a single control function, and is discussed in the following section.

4.2.1.4.1 Control Function 1 - Bulk Sulfur Removal

Either the Stretford or the Giammarco-Vetrocoke process would be suited for the treatment of the slurry drier vent gas and vacuum fractionator off-gas combined with the Flexicoking sour fuel gas. Only the Stretford process for treating the combination of Streams 102 and 153/156 will be discussed here.

4.2.1.4.1.1 Control Technique 1 - Stretford Process

As discussed previously, it might be convenient to combine these two gaseous streams with the Flexicoking sour fuel gas for treatment by the Stretford process. Total combined fuel gas produced would be 55,382 kmol/hr and 27,826 kmol/hr for the base case and MFS case designs, respectively. Prior to Stretford treatment, the combined sour fuel gas from the base case design would contain 0.42% H_2S, 107 ppmv COS and 11.5% CO_2, a composition almost identical to the Flexicoking sour fuel gas. Similarly, the combined sour fuel gas from the MFS case design would be almost identical in composition to the Flexicoking sour fuel gas, and would contain 0.45% H_2S, 90 ppmv COS, and 11.3% CO_2. The treated fuel gas from these combined streams would then contain less than 10 ppmv H_2S and 107 ppmv COS for the base case design, and less than 10 ppmv H_2S and 90 ppmv COS for the MFS case design. Additional sulfur recovered in the Stretford plant would amount to 0.11 Mg/hr for both EDS designs.

The capital investment costs and annualized costs for a single Stretford system designed to handle the combined Streams 304, 102, and 153/156 are presented in Table 4-13. The table also shows the incremental portion of this combined cost that is attributable to Streams 102 and 153/156. Finally, the

TABLE 4-13. COSTS OF BULK SULFUR REMOVAL WITH STRETFORD PROCESS FOR SLURRY DRIER VENT GAS AND VACUUM FRACTIONATOR OFF-GAS (STREAMS 102 and 153/156)

	Base Case		MFS Case	
Cost Element	Total Capital Investment*	Total Annualized Cost†	Total Capital Investment*	Total Annualized Cost†
Stretford Plant for Treating Streams 304, 102 and 153/156				
Total Cost	$45.4 million	$13.9 million	$26.5 million	$7.9 million
% of Base Plant	1.42	1.45	0.92	0.87
Unit Cost	$6.1 million per Mg/hr sulfur recovered	$237/Mg sulfur recovered	$6.7 million per Mg/hr sulfur recovered	$252/Mg sulfur recovered
Incremental Portion of Above Cost Attributable to Streams 102 and 153/156				
Total Cost	$0.41 million	$0.15 million	$0.44 million	$0.16 million
% of Base Plant	0.013	0.016	0.015	0.018
Unit Cost	$3.7 million per Mg/hr sulfur recovered	$175/Mg sulfur recovered	$4.0 million per Mg/hr sulfur recovered	$183/Mg sulfur recovered
Stretford Plant for Treating Streams 102 and 153/156				
Total Cost	$2.3 million	$0.56 million	$2.3 million	$0.56 million
% of Base Plant	0.074	0.058	0.081	0.061
Unit Cost	$21 million per Mg/hr sulfur recovered	$634/Mg sulfur recovered	$21 million per Mg/hr sulfur recovered	$634/Mg sulfur recovered

* The total process cost (TPC) was calculated using: 1) a TPC of $7.8 million for a Stretford unit processing 27.7 Mg/day sulfur (115); and 2) cost increases that vary with the 0.6 power of the capacity ratio. The TPC was converted to total capital investment cost (TCI) using the procedures outlined in Table 4-1. Also, three parallel Stretford units were assumed for the base case design (each processing 59.7 Mg/day sulfur) and two parallel units for the MFS case design (each processing 47.9 Mg/day sulfur), when Stream 304 is combined with Streams 102 and 153/156. Only one Stretford unit is assumed in treating Streams 102 and 153/156 separately (processing 2.66 Mg/day sulfur).

† The annualized costs were derived using the methodology described in Table 4-2 and utilities and chemicals requirements provided in Table A 7-3 of Appendix A-7 (assuming these requirements are linearly proportional to the amount of sulfur processed). Operating labor requirement was assumed to be 0.5 man per shift for combination of Streams 304, 102 and 153/156, and 0.125 man per shift for combination of just Streams 102 and 153/156.

table shows the costs involved in installing and operating a separate Stretford system just for Streams 102 and 153/156. The incentive for combining Streams 102 and 153/156 with Stream 304 is clearly indicated by the much higher costs of the separate Stretford system.

Secondary Waste Streams

Additional Stretford solution purge generated from the treatment of the slurry drier vent gas and vacuum fractionator off-gas would be 33 kg/hr for both EDS designs. Because of the absence of ammonia and mercaptans in the two gas streams, additional contribution to the Stretford oxidizer vent gas would involve only nitrogen, oxygen, and water vapor.

4.2.1.5 Transient Waste Gases (Streams 803, 801, 802)

Transient waste gases are generated from the EDS commercial plant during startups, shutdowns, and upset conditions. In this section, pollution control alternatives applicable to the following three transient waste gas streams are discussed:

- Stream 803 - transient waste gas from liquefaction reactor
- Stream 801 - transient waste gas from Flexicoking
- Stream 802 - transient waste gas from partial oxidation unit (MFS case only).

These transient waste streams are discussed together because the control approach for each is the same, although the streams would not necessarily be combined for treatment. In all likelihood, these transient waste gases would be generated at different times, and pollution control equipment sized to control the largest transient waste gas stream might also be used to handle other transient or intermittent waste gas streams.

The transient waste gas from liquefaction reactor was estimated to be generated at a rate of 3,275 kmol/hr. Transient conditions would be expected twice per year for each of the four reactors, and up to 3 hours in duration each time. This transient waste gas in the Exxon designs was estimated to contain 45.1% hydrocarbons, 0.5% CO, and 2.5% H_2S. The hydrocarbons present include 20 ppmv organics in the 400 to 700°F boiling range. It is likely that some refractory compounds might be present in organics of this boiling range.

The transient waste gas from Flexicoking was estimated to be generated at a rate of 671 kmol/hr and 511 kmol/hr for the base case and MFS case designs, respectively. Transient conditions would be expected 30 times per year for each of the Flexicoking units (three for the base case design and two for the MFS case design), and up to 3 hours in duration each time. For the base case design, this transient waste gas contains 40.9% hydrocarbons, 4.0% CO, and 1.0%

H_2S. For the MFS case design, this transient waste gas contains 39.7% hydrocarbons, 4.0% CO, and 1.5% H_2S. For both designs, the hydrocarbons present include approximately 1% organics with boiling points equal to or higher than that of hexane. This high boiling fraction might contain significant amounts of refractory organic compounds.

The transient waste gas from partial oxidation unit was estimated to be generated at a rate of 12,121 kmol/hr, and applies only to the MFS case design. Transient conditions would be expected once a year, only to one of the partial oxidation units and up to 24 hours in duration. This transient waste gas contains 0.2% methane, 17.7% CO, 0.3% H_2S, 250 ppmv COS, 8 ppmv HCN, and 170 ppmv NH_3.

The approach to control these transient waste gases may involve up to two control functions: incineration and bulk sulfur removal (SO_2 removal after incineration).

4.2.1.5.1 Control Function 1 - Incineration

Incineration of transient waste gases can be accomplished in boilers, dedicated incinerators, or flares. For purposes of discussion, only flares and dedicated thermal incinerators will be considered. Catalytic incineration is not likely to be an attractive alternative, because the presence of sulfur compounds in transient waste gases might interfere or degrade catalysts.

4.2.1.5.1.1 Control Technique 1 - Flaring

Flares are special burners designed to dispose of relatively high volume waste streams consisting of hydrocarbons and other combustibles. Compared to incinerators, flares provide less control of combustion temperature and amount of air feed, but are better equipped to handle sudden surges in gas flow rates and are less expensive. Data on the combustion efficiency of flares are limited. Most recent studies indicate that 96 to greater than 99% of the hydrocarbons in the waste gas are typically destroyed by flaring (Appendix A-15). Hydrocarbon destruction in flares depends on the amount and distribution of oxygen in the combustion zone, and the nature of organic compounds in the gas being burnt. Refractory organic compounds present in the transient waste gases, for example, might experience much lower destruction efficiency than light hydrocarbon gases such as methane or ethane. In addition, most of the CO would be converted to CO_2; NH_3 to nitrogen, NO_x, and water; HCN to nitrogen, NO_x, water, and CO_2; and sulfur compounds to SO_2 and SO_3.

Estimated capital investment costs and annualized costs for flaring of the three transient waste gases in separate systems are presented in Table 4-14. The unit annualized cost ($/kmol waste gas) for flaring of the transient waste gas from Flexicoking is lower, because of the larger flows associated with the longer duration of transient conditions.

4.2.1.5.1.2 Control Technique 2 - Thermal Incineration

Thermal incinerators provide a combustion chamber and mixing between waste gas and air at a high temperature for the oxidation of combustible compounds present in the transient waste gases. Concentrations of CO and hydrocarbons in the incinerated gas would typically range from 10-100 ppmv and 5-20 ppmv, respectively. Assuming combustion with 20% excess air, incineration of the transient waste gas from liquefaction reactor would generate 29,350 kmol/hr of flue gas containing 2,850 ppmv SO_2. For the base case design, incineration of the transient waste gas from Flexicoking would generate 5,610 kmol/hr of flue gas containing 1,250 ppmv SO_2. For the MFS case design, incineration of this same transient waste gas would generate 4,250 kmol/hr of flue gas containing 1,810 ppmv SO_2. Flue gas from incineration of the transient waste gas from partial oxidation unit would be generated at a rate of 23,590 kmol/hr, and contains 1,680 ppmv SO_2.

Estimated capital investment costs and annualized costs for thermal incineration of transient waste gases are presented in Table 4-15. Similar costs are presented in Table 4-16 for thermal incineration with waste heat recovery.

TABLE 4-14. COSTS OF FLARING FOR TRANSIENT WASTE GASES (STREAMS 801, 802, 803)

Cost Element	Stream 803 Transient Waste Gas from Liquefaction Reactor Base Case	 MFS Case	Stream 801 Transient Waste Gas from Flexicoking Base Case	 MFS Case	Stream 802 Transient Waste Gas from Partial Oxidation*
Capital Investment Cost†					
Total Cost, $ million	0.47	0.47	0.096	0.073	1.7
% of Base Plant	0.015	0.016	0.00	0.003	0.060
Unit Cost, $ per kmol/hr waste gas	143	143	143	143	143
Annualized Cost‡					
Total Cost, $ million	0.094	0.094	0.022	0.016	0.34
% of Base Plant	0.0097	0.010	0.002	0.002	0.038
Unit Cost, $/kmol waste gas	1.2	1.2	0.12	0.18	1.2

[*] Generated only from the MFS case design.

[†] The capital investment costs were derived using the equipment cost curve for 200 ft high flares (Figure A 15-3) presented in Appendix A-15 and the methodology outlined in Table 4-1. This capital investment cost corresponds to a unit cost of $143 per kmol/hr waste gas.

[‡] The annualized costs were derived using the methodology described in Section 4.1.3, and rough estimates of manpower and steam requirements.

TABLE 4-15. COSTS OF THERMAL INCINERATION FOR TRANSIENT WASTE GASES (STREAMS 801, 802, 803)

Cost Element	Stream 803 Transient Waste Gas from Liquefaction Reactor		Stream 801 Transient Waste Gas from Flexicoking		Stream 802 Transient Waste Gas from Partial Oxidation*
	Base Case	MFS Case	Base Case	MFS Case	
Capital Investment Cost†					
Total Cost, $ million	3.8	3.8	0.75	0.57	3.1
% of Base Plant	0.12	0.13	0.023	0.020	0.11
Unit Cost, $ per kmol/hr waste gas	1,168	1,168	1,113	1,108	259
Annualized Cost‡					
Total Cost, $ million	0.75	0.75	0.14	0.11	0.62
% of Base Plant	0.078	0.083	0.015	0.012	0.068
Unit Cost, $/kmol waste gas	9.6	9.6	0.78	1.2	2.1

*Generated only from the MFS case design.

†The capital investment costs were obtained using the calculated combustion product flow rates and the installed equipment cost curve (Curve A) presented in Figure A 16-6 of Appendix A-16. This capital investment cost corresponds to a unit cost of $133 per kmol/hr combustion product gas.

‡The annualized costs were derived using the methodology described in Section 4.1.3. Annualized cost for each case is approximately 20% of the corresponding capital investment cost (17.2% capital charge, 2% maintenance, the other costs are small for handling transient waste gases).

TABLE 4-16. COSTS OF THERMAL INCINERATION WITH WASTE HEAT RECOVERY FOR TRANSIENT WASTE GASES (STREAMS 801, 802, 803)

Cost Element	Stream 803 Transient Waste Gas from Liquefaction Reactor		Stream 801 Transient Waste Gas from Flexicoking		Stream 802 Transient Waste Gas from Partial Oxidation[*]
	Base Case	MFS Case	Base Case	MFS Case	
Capital Investment Cost[†]					
Total Cost, $ million	5.0	5.0	0.95	0.72	4.0
% of Base Plant	0.16	0.17	0.030	0.025	0.14
Unit Cost, $ per kmol/hr waste gas	1,526	1,526	1,423	1,416	331
Annualized Cost[‡]					
Total Cost, $ million	0.98	0.98	0.19	0.14	0.79
% of Base Plant	0.10	0.11	0.020	0.016	0.087
Unit Cost, $/kmol waste gas	12.5	12.5	1.0	1.5	2.7

[*] Generated only from the MFS case design.

[†] The capital investment costs were obtained using the calculated combustion product flow rates and the installed equipment cost curve (Curve B) presented in Figure A 16-6 of Appendix A-16. This capital investment cost corresponds to a unit cost of $170 per kmol/hr combustion product gas.

[‡] The annualized costs were derived using the methodology described in Section 4.1.3. Annualized cost for each case is approximately 20% of the corresponding capital investment cost (see footnote for Table 4-15). Steam credits from heat recovery are negligible because these transient waste gases are generated only for very short periods of time.

This latter case would be equivalent to incineration in a boiler. The capital investment costs per unit capacity in Tables 4-15 and 4-16 vary because these unit costs are dependent on the flow rate of the combustion product gas and not the flow rate of the transient waste gas.

4.2.1.5.2 Control Function 2 - Bulk Sulfur Removal

As will be discussed in Section 4.2.2 under Source Type 2 - Combustion Gases, a number of flue gas desulfurization (FGD) processes are commercially available. For control of SO_2 emissions from incinerated transient waste gases, only the sodium throwaway process will be discussed here.

4.2.1.5.2.1 Control Technique 1 - Sodium Throwaway Process

Compared to other FGD systems, the sodium throwaway process is lower in capital investment cost and higher in operating cost. The higher operating cost is the result of the cost of sodium carbonate employed for scrubbing. This higher chemical cost is of lesser concern in the control of intermittent waste streams such as incinerated transient waste gases. Sodium throwaway systems can be designed to remove 90% of the SO_2 present in the inlet stream. Thus, sodium throwaway processes can reduce the SO_2 concentration in the incinerated transient waste gases to 125-285 ppmv.

Estimated capital investment costs and annualized costs for sodium throwaway processes are presented in Table 4-17. Although these costs appear to be relatively high, the use of a common FGD system, sized to control the incinerated transient waste gas from the liquefaction reactor, could also be used to control SO_2 emissions from other transient or intermittent waste gases. There is no need for a dedicated FGD system for each of the transient or intermittent waste gas streams.

TABLE 4-17. COSTS OF SULFUR DIOXIDE CONTROL BY SODIUM THROWAWAY PROCESSES FOR TRANSIENT WASTE GASES (STREAMS 801,802,803)

Cost Element	Stream 803 Transient Waste Gas from Liquefaction Reactor		Stream 801 Transient Waste Gas from Flexicoking		Stream 802 Transient Waste Gas from Partial Oxidation*
	Base Case	MFS Case	Base Case	MFS Case	
Capital Investment Cost†					
Total Cost, $ million	12.2	12.2	2.3	1.8	9.8
% of Base Plant	0.38	0.42	0.073	0.061	0.34
Unit Cost, $ per kmol/hr waste gas	3,722	3,722	3,471	3,454	808
Annualized Cost‡					
Total Cost, $ million	2.5	2.5	0.51	0.39	2.0
% of Base Plant	0.25	0.27	0.053	0.043	0.22
Unit Cost, $/kmol waste gas	31.2	31.2	2.8	4.2	6.7

*Generated only from the MFS case design.

†**Capital investment** costs were obtained using the calculated combustion product flow rates and cost data provided in Reference 45, which indicate a unit capital investment cost of $415 per kmol/hr combustion product gas.

‡The Annualized costs were derived using the methodology described in Section 4.1.3. Sodium carbonate consumption was assumed to be 1.2 mol per mol SO_2 removed. Cost of sodium carbonate was assumed to be $264/Mg.

Note: These cost estimates assume the worst case, that a separate flue gas treatment unit will be installed for each transient stream. Use of a single, common scrubber would reduce these costs.

4.2.1.6 Integrated Control Examples

In this section, examples of combined and sequential control of waste streams are evaluated from the standpoint of the overall emissions reductions achieved and costs incurred. The selection of specific control examples for evaluation in this section is not intended to imply that other technologies could not provide equivalent performance with similar or even lower costs. Specific technologies were selected to cover the types of alternatives which have actually been proposed for facilities in the U.S. Selection of integrated controls will be based upon specific design requirements and local conditions, and can only be made by designers and regulatory authorities involved in a specific project.

The integrated control examples for Source Type 1 address only the combination of three acid gas streams, because this is the one case for which several likely alternatives for multiple controls in series are possible. As discussed in Section 4.2.1.1, it would be logical to combine the following acid gas streams prior to treatment in an integrated control system:

- Stream 508 - acid gas from DEA regenerator
- Stream 501 - acid gas from sour water stripper/ammonia recovery
- Stream 440 - flash gas from partial oxidation unit (MFS case only).

Three integrated control alternatives for the combined acid gas stream will be examined in this section:

- Claus bulk sulfur removal with tail gas treatment by the Beavon Sulfur Removal Process (BSRP).
- Claus bulk sulfur removal with SCOT tail gas treatment, followed by incineration
- Claus bulk sulfur removal with Wellman-Lord tail gas treatment.

The first alternative was proposed for the SRC-I demonstration plant in treating similar acid gas streams, and also by Exxon (although not explicitly stated) in the conceptual design of the EDS commercial plant. The second alternative was proposed for the SRC-II demonstration plant and commercial plant, and the third alternative was proposed in the conceptual design of the H-Coal commercial plant, also for treating similar acid gas streams. Thus, the three control alternatives selected as examples here cover all the cases that have been proposed by the four leading developers of direct liquefaction processes in the U.S.

4.2.1.6.1 Example 1 - Claus Bulk Sulfur Removal with Tail Gas Treatment by the Beavon Sulfur Removal Process

The integrated control system in this example for treating the combined acid gas is illustrated in Figure 4-1. Because of the high concentration level of H_2S in the combined acid gas stream, the Claus plant is operated in the "straight through" mode in this example, with the complete stream sent to the combustion chamber. Ammonia present in the combined acid gas would therefore be mostly combusted to nitrogen and water.

Material flow estimates and performance of the Claus/Beavon integrated control system are presented in Tables 4-18 and 4-19 for the base case and MFS case designs, respectively. The stream names used in these tables are those shown in Figure 4-1. Assumptions used for Claus/Beavon performance included: 1) 95% sulfur removal in the Claus plant; and 2) less than 100 ppmv total sulfur in Beavon off-gas. The Claus tail gas for the base case design is generated at a rate of 3417 kmol/hr, and contains 13,625 ppmv total sulfur (SO_2 equivalent). For the MFS case design, the Claus tail gas is generated at a slightly higher rate of 3926 kmol/hr, because of the presence of CO and hydrogen in the feed gas. The Claus tail gas in this case contains 11,850 ppmv total sulfur (SO_2 equivalent). Total sulfur emissions through the Beavon off-gas were estimated to be less than 0.30 kmol/hr (<10 ppmv H_2S and <90 ppmv COS) for the base case design, and less than 0.35 kmol/hr (<10 ppmv H_2S and <90 ppmv COS) for the MFS case design. These are equivalent to greater than

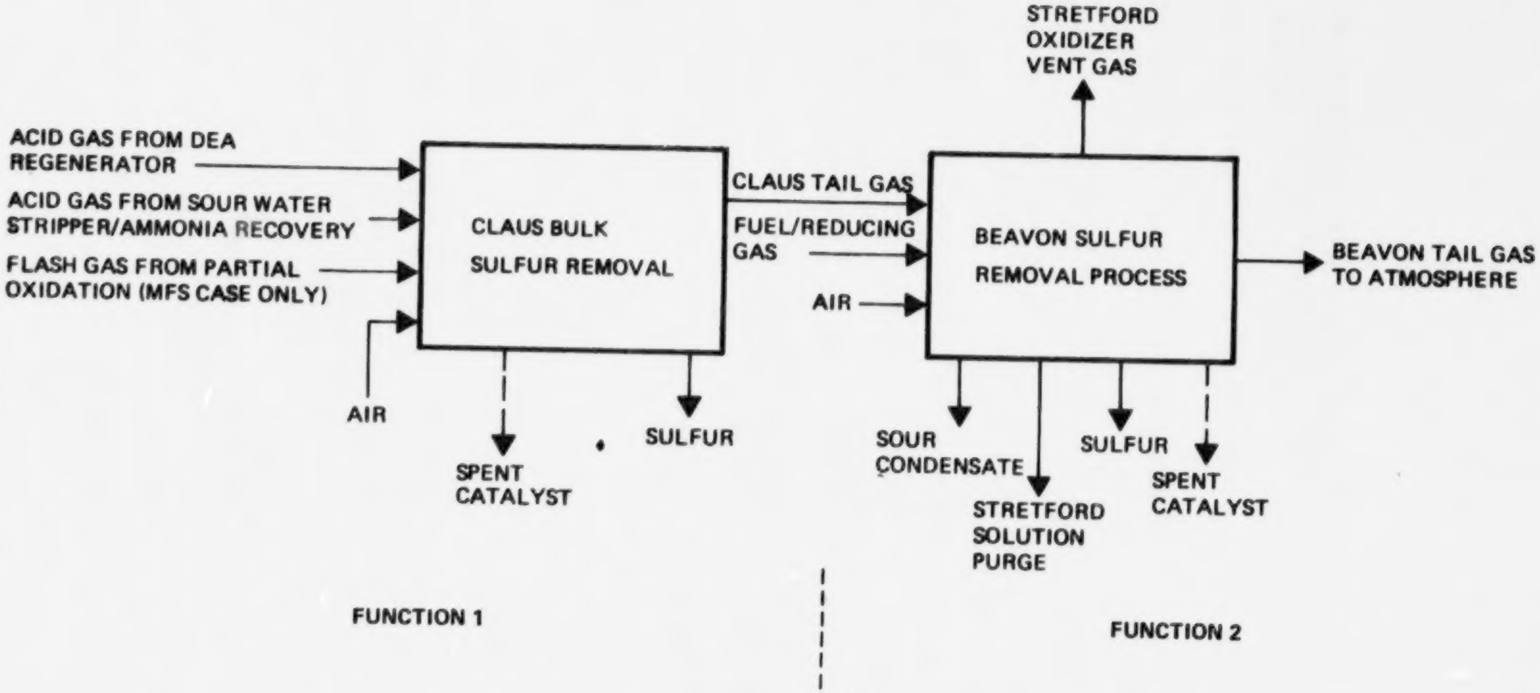
STRETFORD OXIDIZER VENT GAS
ACID GAS FROM DEA REGENERATOR
ACID GAS FROM SOUR WATER STRIPPER/AMMONIA RECOVERY
FLASH GAS FROM PARTIAL OXIDATION (MFS CASE ONLY)
AIR
CLAUS BULK SULFUR REMOVAL
SPENT CATALYST
SULFUR
CLAUS TAIL GAS
FUEL/REDUCING GAS
AIR
BEAVON SULFUR REMOVAL PROCESS
BEAVON TAIL GAS TO ATMOSPHERE
SOUR CONDENSATE
STRETFORD SOLUTION PURGE
SULFUR
SPENT CATALYST
FUNCTION 1
FUNCTION 2

Figure 4-1. Example 1 - Claus bulk sulfur removal with tail gas treatment by Beavon sulfur removal process

TABLE 4-18. MATERIAL FLOW ESTIMATES AND PERFORMANCE OF INTEGRATED CONTROL FOR COMBINED ACID GAS STREAM EMPLOYING CLAUS BULK SULFUR REMOVAL AND BEAVON TAIL GAS TREATMENT (ILLINOIS COAL BASE CASE)

Stream Description/ Control Option	Component	Flow Rate	Concentration
Combined Acid Gas	H_2S	921.7 kmol/hr	62.40 vol %
	NH_3	49.5 kmol/hr	3.35 vol %
	COS	0.45 kmol/hr	0.03 vol %
	Total	1477.2 kmol/hr	
Claus Tail Gas	H_2S	27.7 kmol/hr	8094 ppmv
	SO_2	13.8 kmol/hr	4047 ppmv
	COS	4.2 kmol/hr	1214 ppmv
	CS_2	0.46 kmol/hr	135 ppmv
	Total	3417.2 kmol/hr	
Spent Claus Catalyst	Bauxite/ Alumina	41 Mg/yr	
Sulfur Recovered in Claus Plant	Sulfur	28.1 Mg/hr	
Beavon Off-Gas	H_2S	<0.03 kmol/hr	<10 ppmv
	COS	<0.27 kmol/hr	<90 ppmv
	Total	3033.8 kmol/hr	
Beavon Sour Condensate	Sour Water	17.9 m^3/hr	
Stretford Solution Purge	ADA	0.84 kg/hr	0.19 wt %
	Vanadium	1.68 kg/hr	0.38 wt %
	Na_2CO_3	6.41 kg/hr	1.45 wt %
	$Na_2S_2O_3$	39.5 kg/hr	8.95 wt %
	Na_2SO_4	17.8 kg/hr	4.02 wt %
	H_2O	375.6 kg/hr	85.01 wt %
	Total	441.8 kg/hr	
Spent Beavon Catalyst	Cobalt-Molybdate	20 Mg/yr	
Sulfur Recovered in Beavon Process	Sulfur	1.46 Mg/hr	

TABLE 4-19. MATERIAL FLOW ESTIMATES AND PERFORMANCE OF INTEGRATED CONTROL FOR COMBINED ACID GAS STREAM EMPLOYING CLAUS BULK SULFUR REMOVAL AND BEAVON TAIL GAS TREATMENT (ILLINOIS COAL MFS CASE)

Stream Desciption/ Control Option	Component	Flow Rate	Concentration
Combined Acid Gas	H_2S	920.5 kmol/hr	54.07 vol %
	NH_3	63.8 kmol/hr	3.75 vol %
	COS	0.75 kmol/hr	0.044 vol %
	Total	1702.6 kmol/hr	
Claus Tail Gas	H_2S	27.6 kmol/hr	7040 ppmv
	SO_2	13.8 kmol/hr	3520 ppmv
	COS	4.2 kmol/hr	1057 ppmv
	CS_2	0.46 kmol/hr	117 ppmv
	Total	3926.0 kmol/hr	
Spent Claus Catalyst	Bauxite/ Alumina	47 Mg/yr	
Sulfur Recovered in Claus Plant	Sulfur	28.1 Mg/hr	
Beavon Off-Gas	H_2S	<0.035 kmol/hr	<10 ppmv
	COS	<0.32 kmol/hr	<90 ppmv
	Total	3538.5 kmol/hr	
Beavon Sour Condensate	Sour Water	19.4 m^3/hr	
Stretford Solution Purge	ADA	0.84 kg/hr	0.19 wt %
	Vanadium	1.68 kg/hr	0.38 wt %
	$Na_2\ CO_3$	6.40 kg/hr	1.45 wt %
	$Na_2\ S_2O_3$	39.5 kg/hr	8.95 wt %
	$Na_2\ SO_4$	17.7 kg/hr	4.02 wt %
	H_2O	375.0 kg/hr	85.01 wt %
	Total	441.1 kg/hr	
Spent Beavon Catalyst	Cobalt-Molybdate	20 Mg/yr	
Sulfur Recovered in Beavon Process	Sulfur	1.46 Mg/hr	

99.96% total sulfur removal from the combined acid gas for the Claus/Beavon integrated control system.

There are four secondary waste streams from the Claus/Beavon integrated control system: 1) spent Claus catalyst, 2) Beavon sour condensate, 3) Stretford solution purge, and 4) spent Beavon hydrolysis catalyst. The generation rates of these secondary waste streams are also presented in Tables 4-18 and 4-19. Characteristics of these waste streams and applicable control techniques are described in Sections 4.2.1.1, 4.3, 4.4, and Appendices A-6, A-7 and A-9.

Cost estimates for the Claus/Beavon integrated control system are summarized in Table 4-20. The calculation bases for these cost estimates have been described in Section 4.2.1.1. The capital investment costs for the Claus and the Beavon system are both based on two parallel trains. The capital investment cost for the Claus plant is 28% higher than that for the Beavon plant. However, the total annualized cost for the Claus plant is 65% lower than that for the Beavon plant, primarily as a result of credits for steam generated by the Claus plant. The total capital investment cost and the total annualized cost for the integrated control system represent 1.5 - 1.7% and 1.2 - 1.3% of the respective costs for the uncontrolled EDS base plant. On annualized basis, sulfur can be recovered at a cost of approximately $50/Mg.

4.2.1.6.2 Example 2 - Claus Bulk Sulfur Removal with SCOT Tail Gas Treatment and Incineration

The Claus/SCOT/incineration integrated control system for treating the combined acid gas is illustrated in Figure 4-2. The Claus plant for this example is similar to the Claus plant in Example 1, except the Claus feed also includes a recycle concentrated H_2S stream from the SCOT plant. Incineration of SCOT tail gas is a common practice to minimize odor problems arising from the higher levels of residual H_2S. In addition, incineration is also effective in controlling any residual reduced nitrogen compounds, CO or hydrocarbon present in the SCOT tail gas.

TABLE 4-20. COSTS OF INTEGRATED CONTROL FOR COMBINED ACID GAS STREAM EMPLOYING CLAUS BULK SULFUR REMOVAL AND BEAVON TAIL GAS TREATMENT

	Capital Investment Cost			Annualized Cost		
Control Option	Total Cost, $ Million	% of Base Plant	Unit Cost, $ Million per Mg/hr Sulfur Recovered or Removed	Total Cost, $ Million	% of Base Plant	Unit Cost, $/Mg Sulfur Recovered or Removed
Base Case						
Claus	27.1	0.85	0.97	3.0	0.31	13.6
Beavon	21.1	0.66	14.4	8.5	0.88	735.8
Total	48.2	1.51	1.63*	11.5	1.19	49.3*
MFS Case						
Claus	27.1	0.94	0.97	3.0	0.33	13.6
Beavon	21.1	0.73	14.4	8.6	0.95	747.3
Total	48.2	1.67	1.63*	11.6	1.28	49.9*

*Unit costs for the two control functions are not additive. The total unit cost per Mg/hr sulfur was calculated by dividing the total capital investment cost by total Mg/hr sulfur recovered. Similarly, the total unit cost per Mg sulfur was calculated by dividing the total annualized cost by total Mg sulfur recovered in one year.

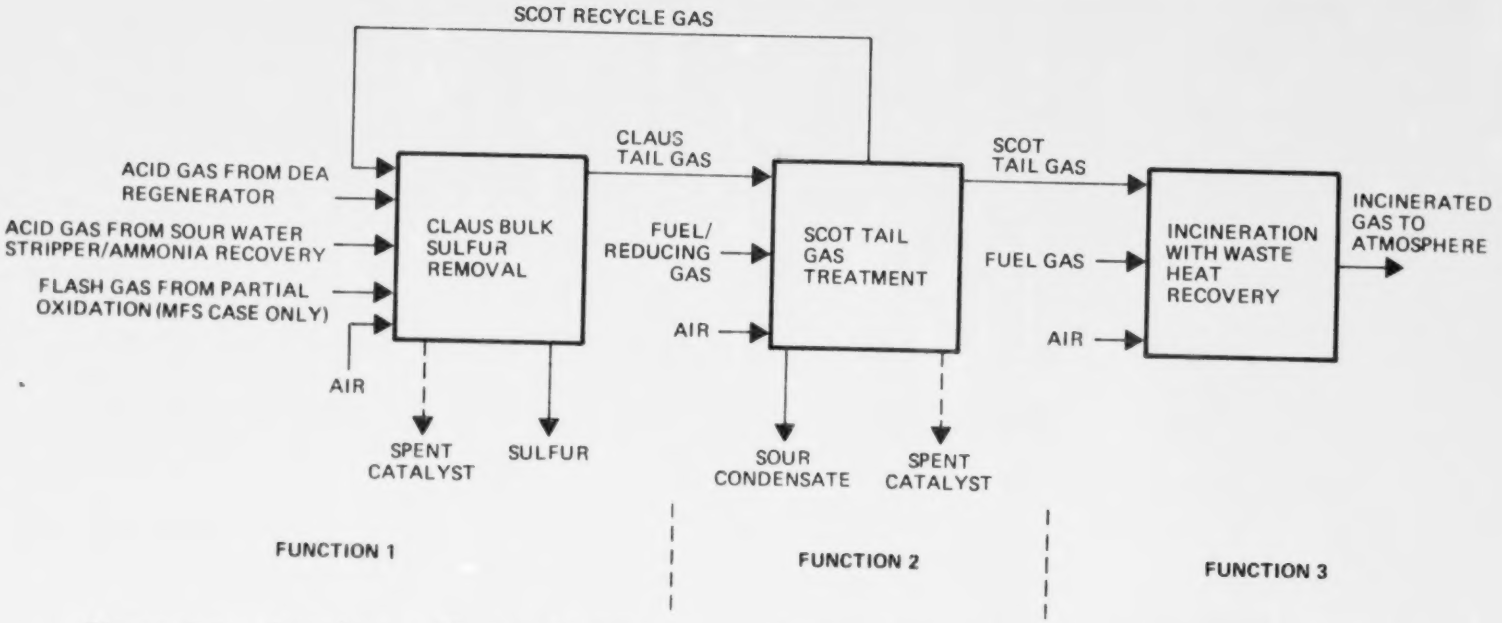
SCOT RECYCLE GAS
ACID GAS FROM DEA REGENERATOR
ACID GAS FROM SOUR WATER STRIPPER/AMMONIA RECOVERY
FLASH GAS FROM PARTIAL OXIDATION (MFS CASE ONLY)
AIR
CLAUS BULK SULFUR REMOVAL
SPENT CATALYST
SULFUR
CLAUS TAIL GAS
FUEL/ REDUCING GAS
AIR
SCOT TAIL GAS TREATMENT
SOUR CONDENSATE
SPENT CATALYST
SCOT TAIL GAS
FUEL GAS
AIR
INCINERATION WITH WASTE HEAT RECOVERY
INCINERATED GAS TO ATMOSPHERE
FUNCTION 1
FUNCTION 2
FUNCTION 3

Figure 4-2. Example 2 - Claus bulk sulfur removal with SCOT tail gas treatment and incineration

Material flow estimates and performance of the Claus/SCOT/incineration integrated control system are presented in Tables 4-21 and 4-22 for the base case and MFS case designs, respectively. The stream names used in these tables are those shown in Figure 4-2. Assumptions used for Claus/SCOT/incineration performance included: 1) 95% sulfur removal in the Claus plant; 2) less than 250 ppmv total sulfur in SCOT off-gas prior to incineration; 3) incineration at 978°K (1300°F) with 20% excess air, with essentially complete conversion of residual H_2S in SCOT off-gas to SO_2. Because of the presence of H_2S in the SCOT recycle gas stream, the Claus tail gases in this example are generated at a higher rate than the Claus tail gases in Example 1. Total sulfur (SO_2 equivalent) levels in the Claus tail gas are 12,920 ppmv and 11,110 ppmv for the base case and MFS case designs, respectively. Total sulfur emissions through the SCOT off-gas were estimated to be less than 0.89 kmol/hr and less than 0.93 kmol/hr for the base case and MFS case designs. These are equivalent to approximately 99.9% total sulfur removal from the Claus/SCOT system. Sulfur emissions (in the form of SO_2) through the incinerated SCOT off-gas are slightly higher because of the presence of sulfur (80 ppmv COS and 3 ppmv H_2S) in the low-Btu fuel gas. Concentration levels of SO_2 in the incinerated SCOT off-gas were estimated to be less than 140 ppmv. Residual CO and organics, if present in the SCOT off-gas, would typically be reduced to concentration levels of 10-100 ppmv and 5-20 ppmv respectively in the incinerated gas (Appendix A-16).

There are three secondary waste streams from the Claus/SCOT/incineration integrated control system: 1) spent Claus catalyst, 2) SCOT sour condensate, and 3) spent SCOT hydrolysis catalyst. The generation rates of these secondary waste streams are also presented in Tables 4-21 and 4-22. Characteristics of these waste streams and applicable control techniques are described in Sections 4.2.1.1, 4.3, 4.4 and Appendices A-6 and A-8.

TABLE 4-21. MATERIAL FLOW ESTIMATES AND PERFORMANCE OF INTEGRATED CONTROL FOR COMBINED ACID GAS STREAM EMPLOYING CLAUS BULK SULFUR REMOVAL WITH SCOT TAIL GAS TREATMENT AND INCINERATION (ILLINOIS COAL BASE CASE)

Stream Description/ Control Option	Component	Flow Rate	Concentration
Combined Acid Gas	H_2S	921.7 kmol/hr	62.40 vol %
	NH_3	49.5 kmol/hr	3.35 vol %
	COS	0.45 kmol/hr	0.03 vol %
	Total	1477.2 kmol/hr	
SCOT Recycle Gas	H_2S	48.3 kmol/hr	16.94 vol %
	CO_2	206.6 kmol/hr	72.46 vol %
	H_2O	30.2 kmol/hr	10.60 vol %
	Total	285.1 kmol/hr	
Claus Tail Gas	H_2S	29.1 kmol/hr	7674 ppmv
	SO_2	14.6 kmol/hr	3838 ppmv
	COS	4.4 kmol/hr	1151 ppmv
	CS_2	0.49 kmol/hr	129 ppmv
	Total	3795.9 kmol/hr	
Spent Claus Catalyst	Bauxite/ Alumina	46 Mg/yr	
Sulfur Recovered in Claus Plant	Sulfur	29.5 Mg/hr	
SCOT Off-Gas	H_2S	<0.63 kmol/hr	<200 ppmv
	COS	<0.16 kmol/hr	<50 ppmv
	Total	3151.6 kmol/hr	
Incinerated SCOT Off-Gas	SO_2	<0.92 kmol/hr	<140 ppmv
	Total	6574.3 kmol/hr	
SCOT Sour Condensate	Sour Water	18.8 m^3/hr	
Spent SCOT Catalyst	Cobalt-Molybdate	12 Mg/yr	

TABLE 4-22. MATERIAL FLOW ESTIMATES AND PERFORMANCE OF INTEGRATED CONTROL FOR COMBINED ACID GAS STREAM EMPLOYING CLAUS BULK SULFUR REMOVAL WITH SCOT TAIL GAS TREATMENT AND INCINERATION (ILLINOIS COAL MFS CASE)

Stream Description/ Control Option	Component	Flow Rate	Concentration
Combined Acid Gas	H_2S	920.5 kmol/hr	54.07 vol %
	NH_3	63.8 kmol/hr	3.75 vol %
	COS	0.75 kmol/hr	0.044 vol %
	Total	1702.6 kmol/hr	
SCOT Recycle Gas	H_2S	48.1 kmol/hr	13.51 vol %
	CO_2	270.2 kmol/hr	75.91 vol %
	H_2O	37.7 kmol/hr	10.58 vol %
	Total	356.0 kmol/hr	
Claus Tail Gas	H_2S	29.1 kmol/hr	6601 ppmv
	CO_2	14.5 kmol/hr	3300 ppmv
	COS	4.4 kmol/hr	990 ppmv
	CS_2	0.48 kmol/hr	109 ppmv
	Total	4405.6 kmol/hr	
Spent Claus Catalyst	Bauxite/ Alumina	53 Mg/yr	
Sulfur Recovered in Claus Plant	Sulfur	29.5 Mg/yr	
SCOT Off-Gas	H_2S	<0.74 kmol/hr	<200 ppmv
	COS	<0.19 kmol/hr	<50 ppmv
	Total	3707.8 kmol/hr	
Incinerated Off-Gas	SO_2	<1.09 kmol/hr	<140 ppmv
	Total	7734.5 kmol/hr	
SCOT Sour Condensate	Sour Water	20.3 m^3/hr	
Spent SCOT Catalyst	Cobalt-Molybdate	12 Mg/yr	

Cost estimates for the Claus/SCOT/incineration integrated control system are summarized in Table 4-23. The calculation bases for these cost estimates have been described in Section 4.2.1.1. However, the costs for the Claus plant are slightly different from those presented in Section 4.2.1.1, because more sulfur is processed in this example as a result of the recycle gas from SCOT. The capital investment costs are based on two parallel trains. Of the total capital investment cost for the integrated control system, approximately 56% are incurred by the Claus plant, 37% by the SCOT plant, and 7% by incineration with waste heat recovery. On annualized cost basis, however, the relative contributions are approximately 18%, 50%, and 32% for Claus, SCOT, and incineration with waste heat recovery. The total capital investment cost and the total annualized cost for the integrated control system represent 1.5 - 1.8% and 1.6 - 1.9% of the respective costs for the uncontrolled EDS base plant. On annualized basis, sulfur can be recovered at a cost of $67-74/Mg.

4.2.1.6.3 Example 3 - Claus Bulk Sulfur Removal with Wellman-Lord Tail Gas Treatment

The Claus/Wellman-Lord integrated control system for treating the combined acid gas is illustrated in Figure 4-3. Unlike the Beavon or SCOT process, incineration of the Claus tail gas prior to SO_2 removal by absorption is an integral part of the Wellman-Lord. Thus, the heating value and the sulfur content of the fuel gas used for incineration have a direct impact on the performance of the Wellman-Lord plant. A concentrated SO_2 stream from the Wellman-Lord regenerator is recycled to the Claus plant for sulfur recovery.

Material flow estimates and performance of the Claus/Wellman-Lord integrated control system are presented in Tables 4-24 and 4-25 for the base case and MFS case designs, respectively. The stream names used in these tables are those shown in Figure 4-3. Assumptions used for Claus/Wellman-Lord performance included: 1) 95% sulfur removal in the Claus plant; and 2) less than 150 ppmv SO_2 in Wellman-Lord off-gas. Since the Wellman-Lord recycle gas contains SO_2,

TABLE 4-23. COSTS OF INTEGRATED CONTROL FOR COMBINED ACID GAS STREAM EMPLOYING CLAUS BULK SULFUR REMOVAL WITH SCOT TAIL GAS TREATMENT AND INCINERATION

	Capital Investment Cost			Annualized Cost		
Control Option	Total Cost, $ Million	% of Base Plant	Unit Cost, $ Million	Total Cost, $ Million	% of Base Plant	Unit Cost, $
Base Case						
Claus	28.0	0.88	0.95 per Mg/hr sulfur recovered	3.0	0.32	13.1/Mg sulfur recovered
SCOT	18.4	0.58	11.9 per Mg/hr sulfur removed	7.9	0.82	648.3/Mg sulfur removed
Incineration with Waste Heat Recovery	3.2	0.10	1.0 per Mmol/hr SCOT tail gas	4.7	0.49	190.0/Mmol SCOT tail gas gas
Total	49.6	1.56	1.68 per Mg/hr sulfur recovered*	15.6	1.63	67.3/Mg sulfur recovered*
MFS Case						
Claus	27.9	0.97	0.94 per Mg/hr sulfur recovered	3.0	0.34	13.1/Mg sulfur recovered
SCOT	18.4	0.64	11.9 per Mg/hr sulfur removed	8.5	0.94	702.5/Mg sulfur removed
Incineration with Waste Heat Recovery	3.5	0.12	0.94 per Mmol/hr SCOT tail gas	5.5	0.61	188.3/Mmol SCOT tail gas
Total	49.8	1.73	1.69 per Mg/hr sulfur recovered*	17.0	1.89	73.4/Mg sulfur recovered*

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[*] Unit costs for the three control functions are not additive. See footnote for Table 4-20.

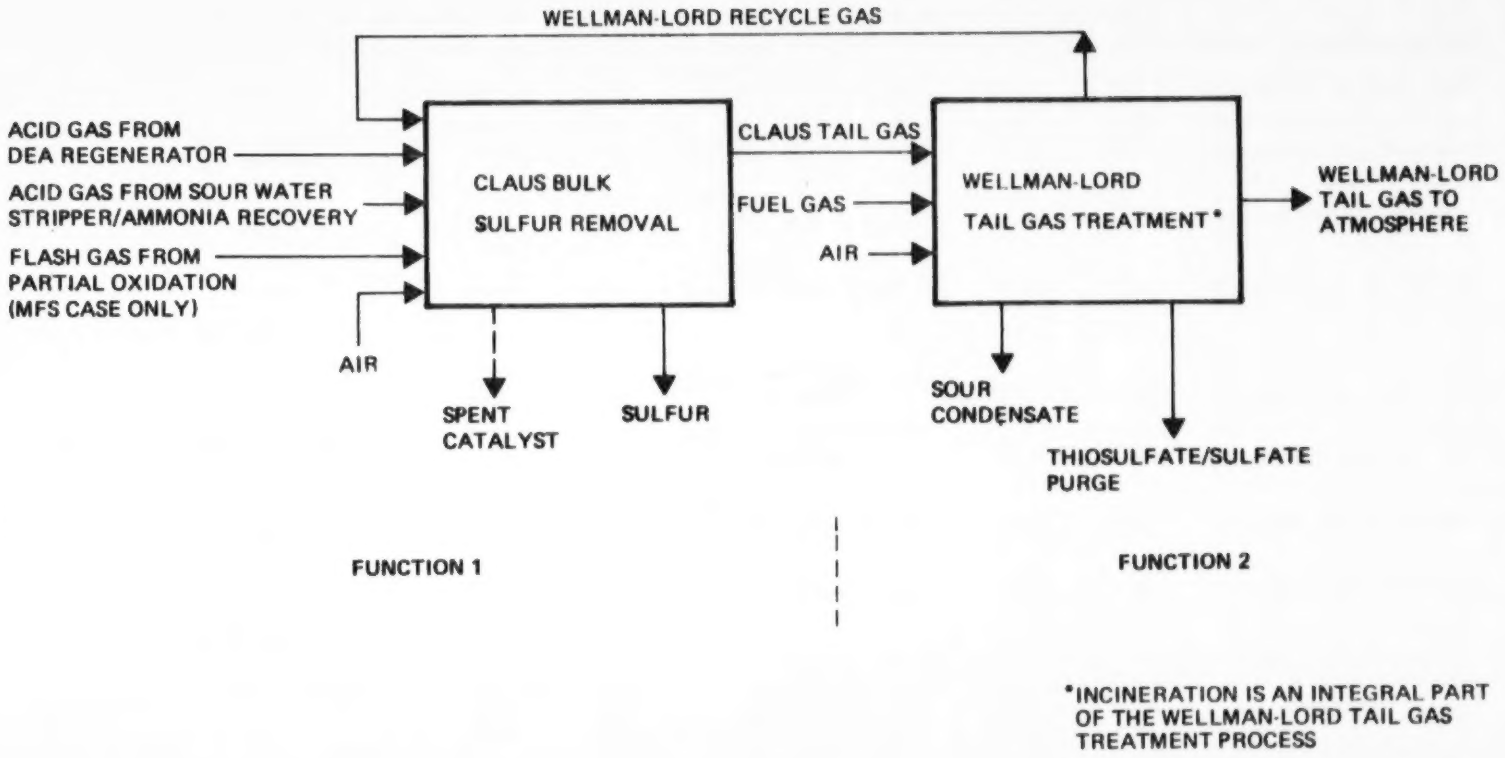
WELLMAN-LORD RECYCLE GAS
ACID GAS FROM DEA REGENERATOR
ACID GAS FROM SOUR WATER STRIPPER/AMMONIA RECOVERY
FLASH GAS FROM PARTIAL OXIDATION (MFS CASE ONLY)
AIR
CLAUS BULK SULFUR REMOVAL
SPENT CATALYST
SULFUR
CLAUS TAIL GAS
FUEL GAS
AIR
WELLMAN-LORD TAIL GAS TREATMENT*
WELLMAN-LORD TAIL GAS TO ATMOSPHERE
SOUR CONDENSATE
THIOSULFATE/SULFATE PURGE
FUNCTION 1
FUNCTION 2
*INCINERATION IS AN INTEGRAL PART OF THE WELLMAN-LORD TAIL GAS TREATMENT PROCESS

Figure 4-3. Example 3 - Claus bulk sulfur removal with Wellman-Lord tail gas treatment

TABLE 4-24. MATERIAL FLOW ESTIMATES AND PERFORMANCE OF INTEGRATED CONTROL FOR COMBINED ACID GAS STREAM EMPLOYING CLAUS BULK SULFUR REMOVAL WITH WELLMAN-LORD TAIL GAS TREATMENT (ILLINOIS COAL BASE CASE)

Stream Description/ Control Option	Component	Flow Rate	Concentration
Combined Acid Gas	H_2S	921.7 kmol/hr	62.40 vol %
	NH_3	49.5 kmol/hr	3.35 vol %
	COS	0.45 kmol/hr	0.03 vol 5
	Total	1477.2 kmol/hr	
Wellman-Lord Recycle Gas	SO_2	45.4 kmol/hr	65.39 vol %
	H_2O	24.0 kmol/hr	34.61 vol %
	Total	69.4 kmol/hr	
Claus Tail Gas	H_2S	28.7 kmol/hr	8796 ppmv
	SO_2	14.4 kmol/hr	4398 ppmv
	COS	4.3 kmol/hr	1319 ppmv
	CS_2	0.48 kmol/hr	147 ppmv
	Total	3267.4 kmol/hr	
Spent Claus Catalyst	Bauxite/ Alumina	39 Mg/yr	
Sulfur Recovered in Claus Plant	Sulfur	29.5 Mg/hr	
Wellman-Lord Off-Gas	SO_2	<1.40 kmol/hr	<150 ppmv
	Total	9322.6 kmol/hr	
Thiosulfate/Sulfate Purge	Na_2SO_3	130.2 kg/hr	41.5 wt %
	$Na_2S_2O_5$	56.8 kg/hr	18.1 wt %
	Na_2SO_4	31.4 kg/hr	10.0 wt %
	$Na_2S_2O_3$	3.4 kg/hr	1.1 wt %
	H_2O	91.8 kg/hr	29.2 wt %
	Total	313.6 kg/hr	
Wellman-Lord Acidic Wastewater	Acidic Wastewater	67.6 m^3/hr	

TABLE 4-25. MATERIAL FLOW ESTIMATES AND PERFORMANCE OF INTEGRATED CONTROL FOR COMBINED ACID GAS STREAM EMPLOYING CLAUS BULK SULFUR REMOVAL WITH WELLMAN-LORD TAIL GAS TREATMENT (ILLINOIS COAL MFS CASE)

Stream Description/ Control Function	Component	Flow Rate	Concentration
Combined Acid Gas	H_2S	920.5 kmol/hr	54.07 vol %
	NH_3	63.8 kmol/hr	3.75 vol %
	COS	0.75 kmol/hr	0.044 vol %
	Total	1702.6 kmol/hr	
Wellman-Lord Recycle Gas	SO_2	45.2 kmol/hr	65.39 vol %
	H_2O	23.9 kmol/hr	34.61 vol %
	Total	69.1 kmol/hr	
Claus Tail Gas	H_2S	28.7 kmol/hr	7598 ppmv
	SO_2	14.4 kmol/hr	3799 ppmv
	COS	4.3 kmol/hr	1141 ppmv
	CS_2	0.48 kmol/hr	127 ppmv
	Total	3777.3 kmol/hr	
Spent Claus Catalyst	Bauxite/ Alumina	46 Mg/yr	
Sulfur Recovered in Claus Plant	Sulfur	29.4 Mg/hr	
Wellman-Lord Off-Gas	SO_2	<1.63 kmol/hr	<150 ppmv
	Total	10,875.8 kmol/hr	
Thiosulfate/Sulfate Purge	Na_2SO_3	130.0 kg/hr	41.5 wt %
	$Na_2S_2O_5$	56.7 kg/hr	18.1 wt %
	Na_2SO_4	31.3 kg/hr	10.0 wt %
	$Na_2S_2O_3$	3.4 kgh/r	1.1 wt %
	H_2O	91.9 kg/hr	29.2 wt %
	Total	313.3 kg/hr	
Wellman-Lord Acidic Wastewater	Acidic Wastewater	76.5 m^3/hr	

less than one-third of the H_2S in the combined acid gas needs to be combusted to SO_2 for subsequent sulfur formation. Thus, less air is needed for combustion, and the Claus tail gases in this example are generated at a lower rate than the Claus tail gases in Examples 1 and 2. Total sulfur (SO_2 equivalent) levels in the Claus tail gas are 14,810 ppmv and 12,790 ppmv for the base case and MFS case designs, respectively. The Claus tail gas is incinerated with a low-Btu fuel gas which has a heating value of 4.7 MJ/Nm^3 (125 Btu/SCF, dry basis) and contains 83 ppmv reduced sulfur compounds. Total sulfur emitted through the Wellman-Lord off-gas was estimated to be less than 1.40 kmol/hr and less than 1.63 kmol/hr for the base case and MFS case designs. These are equivalent to greater than 99.8% total sulfur removal from the Claus/Wellman-Lord integrated control system. Higher sulfur removal efficiencies can be achieved if a higher heating value fuel gas is used for incineration, because total sulfur mass emissions in the Wellman-Lord off-gas (assumed to be less than 150 ppmv) are directly proportional to the flow rate of the off-gas.

There are three secondary waste streams from the Claus/Wellman-Lord integrated control system: 1) spent Claus catalyst, 2) thiosulfate/sulfate purge, and 3) acidic wastewater. The generation rates of these secondary waste streams are presented in Tables 4-24 and 4-25. Characteristics of these waste streams and applicable control techniques are described in Sections 4.2.1.1, 4.3, and 4.4, and Appendices A-6 and A-10.

Cost estimates for the Claus/Wellman-Lord integrated control system are summarized in Table 4-26. The calculation bases for these estimates have been described in Section 4.2.1.1. However, the costs for the Claus plant are slightly different from those presented in Section 4.2.1.1, because more sulfur is processed in this example as a result of the recycle gas from Wellman-Lord. The capital investment costs are based on two parallel Claus units, two parallel Wellman-Lord incinerator-absorber units, and one Wellman-Lord regenerator. The total capital investment cost and the total annualized cost for the integrated control system represent approximately 1.9 - 2.2% and 1.4 - 1.7% of the

TABLE 4-26. COSTS OF INTEGRATED CONTROL FOR COMBINED ACID GAS STREAM EMPLOYING CLAUS BULK SULFUR REMOVAL WITH WELLMAN-LORD TAIL GAS TREATMENT

	Capital Investment Cost			Annualized Cost		
Control Option	Total Cost, \$ Million	% of Base Plant	Unit Cost, \$ Million per Mg/hr Sulfur Recovered or Removed	Total Cost, \$ Million	% of Base Plant	Unit Cost, \$/Mg Sulfur Recovered or Removed
Base Case						
Claus	27.9	0.88	0.95	3.0	0.32	13.1
Wellman-Lord	33.8	1.06	22.2	11.1	1.15	926.8
Total	61.7	1.94	2.09*	14.1	1.47	60.9*
MFS Case						
Claus	27.9	0.97	0.95	3.0	0.34	13.1
Wellman-Lord	33.8	1.17	22.4	11.8	1.30	990.0
Total	61.7	2.14	2.10*	14.8	1.64	63.9*

*Unit costs for the two control functions are not additive. See footnote for Table 4-20.

respective costs for the uncontrolled EDS base plant. On annualized basis, sulfur can be recovered at a cost of $61-64/Mg with the Claus/Wellman-Lord integrated control system.

4.2.2 Source Type 2 - Combustion Gases

There are five gaseous waste streams under Source Type 2 for the base case design, and six gaseous waste streams under Source Type 2 for the MFS case design. These waste streams are:

- Stream 707a - flue gas from power generation system
- Stream 701a - flue gas from steam generation system
- Stream 107 - flue gas from liquefaction slurry preheat furnace
- Stream 161 - flue gas from partial oxidation feed vacuum fractionator preheat furnace (MFS case only)
- Stream 203 - flue gas from solvent hydrogenation fuel preheat furnaces
- Stream 434 - flue gas from hydrogen plant reformer furnaces (base case only)
- Streams 449/450 - regeneration/decommissioning off-gas from high and low temperature shift catalysts (MFS case only).

With the exception of Streams 449/450, these gaseous waste streams are generated by the combustion of coal and/or fuel gas. Streams 449/450 are included under this source type because the characteristics of the combined stream are very similar to those of flue gases from fuel combustion.

Combustion of fossil fuels produces a flue gas stream that contains SO_2, NO_x, and/or particulates. Sulfur dioxide is formed rapidly in the combustion process when sulfur contained in the fuel reacts with oxygen in the air. Variations in the combustion process are not effective in reducing SO_2 emissions. Rather, sulfur must be removed from the fuel or, once formed, SO_2 must be removed from the exhaust gas.

The generation of NO_x from air-fed fuel combustion processes occurs by two separate mechanisms, namely thermal NO_x formation and fuel NO_x formation. Thermal NO_x results from the thermal fixation of molecular nitrogen and oxygen in the combustion air and is sensitive to flame temperatures and to local concentrations of oxygen. Fuel NO_x is created from the oxidation of organically-bound nitrogen in the fuel being combusted. Fuel NO_x formation is strongly affected by the rate of mixing of the fuel and air and by the local oxygen concentration. Approximately 95% of oxides of nitrogen from combustion are emitted as NO, about 5% are emitted as NO_2, and trace quantities are emitted as other nitrogen oxides.

The particulates generated during combustion result mainly from the ash content of the fuel. The partition of ash emissions between fly ash in the flue gas and bottom ash is a function of combustion unit design. Particulate emissions from combustion of gaseous or oil-based fuels are significantly less than for combustion of coal.

Approaches to treatment of gaseous waste streams under Source Type 2 may involve up to three control functions:

1) NO_x control
2) Particulate removal
3) SO_2 removal

NO_x Control

NO_x pollution control techniques are of two types: 1) those that limit nitrogen conversion to NO_x by modifying combustion characteristics and 2) flue gas treatment techniques.

Combustion modification techniques are the most widely used techniques. They can achieve from 25 to 60% reduction in NO_x emissions. Some of the common combustion modification techniques are: 1) low excess air, 2) staged combustion, 3) flue gas recirculation, 4) reduced load, 5) low NO_x burners, and 6) ammonia injection. The key features and unit costs of these techniques are discussed in Table 4-27.

Low excess air level in the furnace has generally been found to be an effective method for NO_x control. In this technique, the combustion air is reduced to the minimum amount required for complete combustion, while maintaining acceptable furnace cleanliness, and steam temperature. With less

TABLE 4-27. COMBUSTION MODIFICATION TECHNIQUES FOR NO_x CONTROL

Control Technique	Description of Technique	Efficiency (as % NO_x Reduction)	Type of Fuel Fired	Range of Application
Low Excess Air (LEA)	Combustion air is reduced to the minimum amount required for complete combustion while maintaining proper stream temperature.	0 - 25	Pulverized coal	Excess O_2 lowered to 5.2% on the average.
		5 - 25	Stoker coal	Excess O_2 limited to 5-6% minimum.
		0 - 28 0 - 24	Residual oil Distillate oil	Excess O_2 can be reduced to <3%.
		5 - 35	Natural gas	Excess O_2 can be reduced to <3%.
Staged Combustion				
Overfire Air Injection (OFA)	Injection of air above the top burner level through OFA ports together with a reduction in air flow to the burners (staged combustion).	5 - 30	Pulverized coal	Burner O_2 can be as low as stoichiometric
Staged Combustion Air (LEA + OFA)	Reduction of under grate air flow and increase of overfire air flow.	5 - 25	Stoker coal	Excess O_2 limited to 5% minimum.
	Fuel rich firing burners with secondary combustion air ports.	20 - 50 17 - 44	Residual oil Distillate oil	70-90% burner stoichiometry can be used with proper burner installation.
	Injection of secondary air downstream of the burner(s) in the direction of the flue gas path.	5 - 46	Natural gas	70-90% burner stoichiometries can be maintained. Applicable to all units, however, requires extensive equipment modification.

(Continued)

TABLE 4-27. (CONTINUED)

Control Technique	Stage of Development	Secondary Waste	Cost	Limitations and Comments
Low Excess Air (LEA)	Available but implemented on a limited basis only	None	Capital:* \$440 to \$550/MW heat input Operating: 0 to 8 mills/10^3 kg steam	Limited by increase in CO, HC and particulate emissions. Increase in boiler efficiency may be achieved as a benefit
	Available now but need R&D on lower limit of air	None	Capital: \$600 to \$1850/MW heat input Operating: 13 to 57 mills/10^3 kg steam	Danger of overheating grate, clinker formation, corrosion and high CO emissions
	Available	None	Capital: \$460 to \$2400/MW of heat input Operating: <83 mills/10^3 kg steam	Added benefits included increase in boiler efficiency. Limited by increase in CO, HC and TSP emissions.
	Available	None	Capital: \$580/MW of heat input Operating: --	Generally practical because of increase boiler efficiency. Best NO_x reductions reported for large multi-burner units.
Staged Combustion				
Overfire Air Injection (OFA)	Available but not demonstrated	None	Capital: \$800 to \$940/MW heat input Operating: 80 to 85 mills/10^3 kg steam	Limited by possible increase in slagging and corrosion. Excess air may be required to ensure complete combustion, thereby decreasing efficiency.
Staged Combustion Air (LEA + OFA)	Most stokers have OFA ports as smoke control but may need better air flow control devices	None	Capital: \$600 to \$800/MW heat input Operating: 24 to 32 mills/10^3 kg steam	Overheating grate, corrosion and high CO emissions can occur if under grate air flow is reduced below acceptable level as in LEA
	Technique is applicable on packaged and field-erected units. However, not commercially available for all design types.	None	Capital: \$870 to \$5150/MW of heat input Operating: 123 to 370 mills/10^3 kg steam	Best implemented on new units. Retrofit is probably not feasible for most units, especially packaged ones.
	Technique is still experimental especially for small firetube and watertube units	None	Capital: \$1070/MW of heat input Operating: 117 mills/10^3 kg steam	Found to be less effective on firetube boilers than watertube boilers. Generally less effective for gas-fired units.

(Continued)

* Capital costs presented in this table are total process costs (TPC) and do not include interest during construction.

TABLE 4-27. (CONTINUED)

Control Technique	Description of Technique	Efficiency (as % NO_x Reduction)	Type of Fuel Fired	Range of Application
Staged Combustion				
Air and Fuel Rich Firing	One or more burners fired on air only. Remainder of burners firing fuel rich.	27 - 39	Pulverized coal	Boilers must have a minimum of 4 burners, or designed with excess burners.
Flue Gas Recirculation (FGR)	Recirculation of the flue gas to the burner windbox.	0 - 20	Pulverized coal	A maximum of 25% of the flue gas can be recirculated.
		15 - 30 58 - 73	Residual oil Distillate oil	Up to 25-30% of flue gas recycled. Can be implemented on all design types.
		48 - 86	Natural gas	Flue gas recirculation rates possible up to 45%. Technique is applicable to all boiler types except ones equipped with ring burners.
Reduced Load	Reduction of fuel and air flow to the burner.	Up to 45%	Pulverized coal	Load may be reduced to 25% of capacity.
		Average 15%	Stoker coal	Load may be reduced to 25%.

(Continued)

TABLE 4-27. (CONTINUED)

Control Technique	Stage of Development	Secondary Waste	Costs	Limitations and Comments
Staged Combustion				
Air and Fuel Rich Firing	Available, but engineering refinement is needed prior to implementation	None	Not available	Load reduction required in most cases. Possible increased slagging and corrosion. New boiler design will incorporate the required number of burners.
Flue Gas Recirculation	Not offered because the method is comparatively ineffective	None	Not available	Flue gas recirculation lowers the bulk furnace gas temperatures and reduces O_2 concentration in the combustion zone. Requires installation of flue gas recirculation ducts, fans, etc. Many cause combustion instability.
	Available. Requires extensive modifications to the burner and windbox	None	Capital: $1070 to $5150/MW of heat input Operating: 196 to 438 mills/10^3 kg of steam	Best wuited for new units. Costly to retrofit. Possible flame instability at high FGR rates.
	Available now. Best suited for new boilers. Retrofit application would result in extensive burner modifications.	None	Capital: $870 to $1070/MW of heat input	Flame instability problem is not severe except for ring burners. Minor burner modifications can guarantee stable flames. Most effective on watertube units.
Reduced Load	Available, but not implemented because of negative operational impacts	None	Not available	Best used with increase in firebox size for new boilers. Load reduction may not be effective because of increase in excess O_2.
	Available	None	Not available	Only for stokers that can reduce load without increasing excess air. Not a desirable technique because of loss in boiler efficiency.

(Continued)

TABLE 4-27. (CONTINUED)

Control Technique	Description of Technique	Efficiency (as % NO_x Reduction)	Type of Fuel Fired	Range of Application
Reduced Load		33% decrease to 25% increase*	Residual oil	Applicable to all boiler types and sizes. Load can be reduced to 25% of maximum
		31% decrease to 17% increase*	Distillate oil	
		32% decrease to 82% increase*	Natural gas	Tests to 20% of rated capacity. Applicable to all units.
Low NO_x Burners	New burner designed to utilize controlled air-fuel mixture	45 - 60	Pulverized coal	Prototypes are limited to size ranges >30 MW
		20 - 50	Residual oil	New burners described generally applicable to all boilers
		20 - 50	Distillate oil all boilers.	
		20 - 50	Natural gas	More specific information needed
NH_3 Injection	Injection of NH_3 into convective section of the boiler	40 - 60	Pulverized coal	NH_3 injection rate limited to $\frac{NH_3}{NO} = 1.5$
		40 - 70	Residual oil	Applicable for large package and field-erected watertube boilers
		40 - 70	Distillate oil	
		40 - 70	Natural gas	Not feasible for firetube boilers

(Continued)

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* Increase in NO_x is a surprising result indicated by limited test data, possibly due to higher air/fuel ratio under reduced load conditions. These results will need confirmation.

TABLE 4-27. (CONTINUED)

Control Technique	Type of Fuel Fired	Stage of Development	Secondary Waste	Costs	Limitations and Comments
Reduced Load	Residual Oil Distillate Oil	Available now as a retrofit application. Better implementation with improved firebox design.	None	Not Available	Technique not effective when it necessitates an increase in excess O_2 levels. RL is possible to implement in new designs as reduced combustion intensity [enlarged furnace plan area].
	Natural Gas	Technique available. However, retrofit application is not feasible due to initial low load factor of industrial units.	None	Not Available	Least effective on firetube boilers because of lower combustion intensity. Applicable for new watertube units with increased firebox size.
Low NO_x Burners	Pulverized Coal	Development stage prototypes are available from major boiler manufacturers.	None	Capital: $800 to $940/MW of heat input Operating: 80 to 85 mills/10^3 kg steam	Low NO_x burners could maintain the furnace in an oxidizing environment to minimize slagging and reduce the potential for furnace corrosion. More complete carbon utilization results because of better coal/air mixing in the furnace. Lower O_2 requirements may be obtained with all the combustion air admitted through the burners.
	Residual Oil Distillate Oil	Commercially offered but not demonstrated.	None	Capital: $860 to $5150/MW of heat input Operating: 123 to 376 mills/10^3 kg steam	Specific emissions data from oil fired industrial boilers equipped with LNB are lacking.
	Natural Gas	Commercially offered but not demonstrated.	None	Capital: $860 to $1070/MW of heat input Operating: 106 to 117 mills/10^3 kg steam	Specific emissions data from gas fired industrial boilers equipped with LNB are lacking.
NH_3 Injection	Pulverized Coal	Commercially offered but not demonstrated.	Ammonium Bisulfate	Capital: $4800/MW of heat input Operating: 247 mills/10^3 kg steam	Limited by furnace geometry. Performance is sensitive to flue gas temperature and residence time at optimum temperatures. By-product emissions such as ammonium bisulfate could cause operational problems.
	Residual Oil Distillate Oil	Commercially offered but not demonstrated.	Ammonium Bisulfate	Capital: $4940 to $9770/MW of heat input Operating: 266 to 433 mills/10^3 kg of steam Not available for natural gas	Some increased maintenance of air heater/economizer parts might be necessary when burning high sulfur oil. Technique is very costly. Should have fewer problems when firing natural gas.
	Natural Gas	Not available.	---	---	---

Data source: Reference 46 and 47.

oxygen available in the flame zone, both thermal and fuel NO_x formation are reduced. In addition, the reduced air flow lowers the quantity of flue gas released resulting in an improvement in boiler efficiency.

Staged combustion produces overall fuel-rich conditions during the first couple of seconds and promotes the reduction of NO to N_2. Various methods to achieve this are available. Overfire Air and Burners Out of Service are two techniques generally used on coal fired boilers. Details regarding their performance and applicability are provided in Table 4-27.

Flue gas recirculation consists of recycling a portion of the flue gas back to the primary combustion zone. This reduces NO formation by lowering the bulk gas temperature and oxygen concentration. This technique, however, is effective only on oil and gas fired boilers.

Load reduction can be used to decrease NO_x emissions. Thermal NO_x formation generally increases as the volumetric heat release rate or combustion intensity increases. Reduced combustion intensity can be brough about by load reduction by either derating the boiler or using an enlarge firebox.

Low NO_x burners have been developed primarily for reducing NO_x emissions from utility boilers. Their principal characteristics are reduced flame turbulance, delayed fuel air mixing, and establishment of fuel-rich zones where combustion initially takes place. It is now standard practice for all utility boilers to come equipped with low NO_x burners.

The process of injecting ammonia was developed by Exxon Research and Engineering Company under the trade name of Thermal $DeNO_x$. This technique acts by reducing NO to elemental nitrogen and oxygen with NH_3 at flue gas temperatures ranging from approximately 1070 to 1290°K. However, optimal NO reduction occurs over a very narrow temperature range, around 1240 ± 50°K.

The cost of combustion modification techniques for controlling NO_x emissions depend upon: 1) the additional hardware required, and 2) any changes in operational procedures that may increase the cost of steam production. Cost estimates for combustion modification techniques are provided in Table 4-27.

Flue gas treatment techniques have been proposed for control of NO_x emissions to levels significantly below those achievable by combustion modification techniques. Although large scale flue gas treatment schemes have not been proved commercially in the U.S., these techniques are being applied in Japan. The key features of some of these processes are provided in Table 4-28.

Of the processes listed in Table 4-28, the Selective Catalytic Reduction (SCR) system, using ammonia to react with NO_x is perhaps the most promising. SCR commercial units are being applied on many gas and oil-fired boilers in Japan. Full scale tests of coal-fired boilers are scheduled in mid-1982; two units are operating and several more are due online. Considerable advances have been made; however, some significant technical/economic questions must be answered before widespread application of SCR units can occur. At this time, flue-gas treatment techniques are not commercially utilized in the U.S. Control techniques for nitrogen oxides are further discussed in Appendix A-21.

Particulate Removal

The choice of the particulate collection equipment depends upon a number of factors: the properties of the particulate such as particle size distribution and resistivity; the concentration and volume of the particulate to be handled; the temperature, humidity and chemical composition of the gaseous medium; and most importantly the collection efficiency required.

There are four basic types of particulate collection equipment: 1) cyclones, 2) fabric filters/baghouses, 3) venturi scrubbers and 4) electrostatic precipitators. The key features and unit costs of these collection devices are presented in Table 4-29. In addition, there are other particulate control devices which use a combination of the above approaches. However, these are not addressed here.

TABLE 4-28. NO_x FLUE GAS TREATMENT CONTROL ALTERNATIVES FOR BOILERS

Control Technique	Description of Technique	Principle of Operation	Efficiency (% as NO_x Reduction)	Applicability
Selective Catalytic Reduction (SCR) - Fixed Packed Bed Reactors	Utilizes NH_3 to selectively reduce NO_x to N_2.	Reactor contains ring shaped catalyst pellets packed in fixed bed.	Up to 90%	Applicable only to flue gas streams containing particulate emissions of less than 20 mg/Nm^3.
- Moving Bed Reactors	Utilizes NH_3 to selectively reduce NO_x to N_2.	Reactor contains catalyst (rings or pellets) gravity-fed mechanically-screened, and returned to reactor.	Up to 90%	Applicable only to flue gas streams containing particulates at less than 1 g/Nm^3.
- Parallel Flow Reactor	Utilizes NH_3 to selectively reduce NO_x to N_2.	Reactor contains a special catalyst arrangement (honeycomb, parallel plate, or tubes).	Up to 90%	Testing currently under way for high particulate (>1 g/Nm^3) flue gas.
Absorption-Oxidation	Removes NO_x from flue gas by absorbing the NO or NO_x into a solution containing an oxidant which converts the NO_x to a nitrate salt.	Use of gas/liquid contactors. Perforated plate and packed towers accomplish NO_x absorption by generating high gas/liquid ratio.	The relative insolubility of NO in water will prohibit a high efficiency.	No published information available.
SCR - NO_x/SO_x Removal	Utilizes NH_3 to catalytically reduce NO_x after SO_x is absorbed and reacted with catalyst.	Reactor and catalyst removes both NO_x and SO_2, uniquely designed parallel flow reactor used to avoid particulate problems.	80% NO_x reduction, 90% SO_x reduction (theoretical).	Should be applicable to high particulate flue gas.

(Continued)

TABLE 4-28. (CONTINUED)

Control Technique	Stage of Development	Secondary Wastes	Costs	Limitations and Comments
SCR - Fixed Packed Bed Reactors	Commercially available only for natural gas-fired boilers at this time.	Spent catalyst	20MW estimate: Capital: * $155/kW (coal) $84/kW (oil) $32/kW (gas) Operating: 2.5 mills/kWh (coal) 2.3 mills/kWh (oil) 1.4 mills/kWh (gas)	Although it is possible to install a hot ESP to reduce the particulate level to 20 mg/Nm^3, this is expensive and not always effective. Thus, fixed bed SCR systems are not considered for application to coal-fired boilers.
SCR - Moving Bed Reactors	Has been applied in Japan to several oil-fired industrial and utility boilers.	Spent catalyst	20MW estimate: Capital: $110/kW (coal) $84/kW (oil) Operating: 2.4 mills/kWh (coal) 2.5 mills/kWh (oil)	Although it is possible to install a hot ESP to reduce the particulate level to 1 g/Nm^3, this is expensive and not always effective. Thus, moving bed SCR systems are not considered for application to coal-fired boilers.
SCR - Parallel Flow Reactor	Had been applied in Japan to several oil-fired industrial and utility boilers. Applicability to coal-fired boilers currently being tested by EPA.	Spent catalyst	20MW estimate: Capital: $53/kW (coal) $46/kW (oil) Operating: 1.8 mills/kWh (coal)	Greatly reduces particulate impaction as gas flow is parallel to catalyst surface. Unreacted NH_3 downstream can react with SO_2 or SO_3 to form ammonium bisulfate or the NH_3 could enter FGD and ESP equipment.
Absorption-Oxidation	No coal-fired tests have been made.	NO_3^- salts in wastewater	20MW estimate Capital: $597/kW (coal Operating: 9.6 mills/kWh (coal)	The presence of particulates requires a prescrubber. The presence of SO_2 requires FGD pretreatment. Increased NO_x concentration requires a larger column height and increased oxidant concentration. Nitrate salts formed as a secondary pollutant.
SCR - NO_x/SO_x Removal	No continuous coal-fired NO_x removal test data for NO_x/SO_x systems are available.	Spent catalyst	20MW estimates Capital: $567/kW (coal & oil) Operating: 6 mills/kWh (coal & oil)	System is not affected by changes in the boiler gas flow rate or particulate concentrations. Changes in NO_x concentration because of boiler load changes may be compensated for by conventional control system used with the NH_3 injection equipment.

(Continued)

* Capital costs presented in this table are total process costs (TPC) and do not include interest during construction.

TABLE 4-28. (CONTINUED)

Control Technique	Description of Technique	Principle of Operation	Efficiency (% as NO_x Reduction)	Applicability
Adsorption NO_x/SO_x Removal	Adsorbed NO_x is reduced to N_2 while SO_2 is reduced and condensed to elemental S	The adsorption process removes NO_x and SO_2 from flue gas by absorbing them onto a special activated char	40-60% NO_x reduction, 80-95% SO_x reduction	May be applicable to handle coal
Electron Beam Radiation NO_x/SO_x Removal	A dry process that utilizes an electron beam to bombard the flue gas, thereby removing NO_x and SO_2	Flue gas is taken from the boiler air preheater and passed through a cold ESP tp remove particulate. NH_3 is added and the gas is then bombarded with an electron beam.	Removal efficiency will decrease as NO_x and SO_2 increase	By-product treatment technology needs to be more fully developed before commercialization
Adsorption-Reduction NO_x/SO_x Removal	Simultaneously removes NO_x and SO_2 from flue gas by absorbing them into a scrubbing solution	Based on the use of Chelating compounds complexed with iron to "catalyze" the absorption of NO_x	60-70% NO_x reduction, 90% SO_2 reduction	Applicable only to high sulfur coals
Oxidation-Absorption-Reduction NO_x/SO_x	Simultaneously removes NO_x and SO_2 from flue gas by oxidizing NO to NO_2 and then absorbing NO_2 and SO_2 into a scrubbing solution	Based on the use of gas-phase oxidants, either O_3 or ClO_2, to selectively oxidize NO to NO_2	90% NO_x reduction, 95% SO_2 reduction for oil-fired tests	Not applicable to low sulfur coals
Oxidation-Absorption	Excess O_3 is used to selectively oxidize NO_x to N_2O_5	N_2O_5 formed by oxidation is absorbed into aqueous solution and concentrated to form a 60% HNO_3 by-product	Not available	May be applicable to handle high particulate flue gas

(Continued)

TABLE 4-28. (CONTINUED)

Control Technique	Stage of Development	Secondary Wastes	Costs	Limitations and Comments
Adsorption NO_x/SO_x Removal	Presently in the prototype unit stage of development.	Ash for disposal	20MW estimates: Capital: $257/kW (coal) Operating: 2.7 mills/kWh (coal)	Very complex process. Numerous process steps involve hot solids handling with numerous mechanical problems possible.
Electron Beam Radiation NO_x/SO_x Removal	No coal-fired tests have been performed at this time.	Ammonium nitrates and sulfates	20MW estimates: Capital: $241/kW (coal) Operating: N/A (coal)	NO_x/SO_2 removal will drop off drastically at low radiation doses based on oil-fired pilot tests. Sulfate and nitrate salts as well as other ionic species formed as by-products.
Adsorption-Reduction NO_x/SO_x Removal	Preliminary testing stage of development.	Sulfate and nitrate salts and gypsum	20MW estimates: Capital: $493/kW (coal) $223/kW (oil) Operating: 8.8 mills/kWh (coal) 6.4 mills/kWh (oil)	Requires large absorbers with high liquid rates. Absorbing solution is highly corrosive; and sulfate and nitrate salts formed as secondary pollutants.
Oxidation-Absorption-Reduction NO_x/SO_x	Prototype stage of development. No coal-fired flue gas tests have been performed at this time.	NO_3^- or N-S salts or NH_3 based compounds in wastewater	20MW estimates: Capital: $278/kW (oil) Operating: 7.6 mills/kWh	Costly gas-phase oxidants create secondary wastewater pollution problems. The use of ClO_2 introduces a chloride pollutant problem.
Oxidation-Absorption	One coal-fired test has been performed with no published information.	NO_3^- salts in wastewater gypsum	Not available	Production of nitrate salts poses a potential secondary pollution problem. Corrosion problems.

Data source: Reference 48.

TABLE 4-29. KEY FEATURES OF PARTICULATE COLLECTION EQUIPMENT

Control Device	Operating Principle	Removal Efficiency Range (weight %)	Inlet Loading Limitation (g/Nm^3)	Normal Pressure Drop Range (cm H_2O)	Reliability or Other Limitations	Secondary Waste	Capital Costs*	General Comments
Cyclone	Particulates removed from gas stream by imparting a centrifugal force. The inertia of the particulates carries them to the walls where they fall to the bottom of the cyclone for removal.	50 to 80% for >5 µm. 80 to 95% for 5 to 20 µm.	>2.4	7 to 20	Cannot effectively remove particulates smaller than 5 µm.	Collected particulates	About \$212/m^3/min for total installed system. Costs vary with particle loading and size distribution, and collection efficiency specified.	High reliability due to simple operating principle with no moving parts. Low energy requirements.
Fabric Filter (Baghouse)	Fabric filter material is arranged in a tubular shape with the particulate laden gas stream passing through the filter. Particulate removal primarily results from the buildup of collected material on the dirty-air side of the filter. The filter is periodically cleaned by mechanical shaking or a pressurized reverse air flow, or by a pulsed jet.	98.5 to 99.5% for 0.25 to 0.5 µm. 99 to 99.5% for 0.75 to 1 µm. 99.5 to 99.9% for 3 µm. 99.95% for 3 µm.	>0.24	5 to 25	Plugging problems will result if condensation occurs on filter media or if hygroscooic-material is collected. Temperature limit varies with type of filter media used, maximum is 560 K.	Collected particulates	About \$282/m^3/min for total system. Costs vary with particle loading and size distribution, gas stream temperature, fabric material, mode of cleaning, and collection efficiency specified.	High particulate removal efficiency. High instaliation cost. Large scale required.
Venturi Scrubber	Removal of particulates from a gas stream by intimate concontact with scrubbing water and droplets. Agglomerated particulates are subsequently removed in a centrifugal and/or mist eliminator.	50 to 92.5% for 0.25 µm. 60 to 98% for 0.5 µm. 70 to 99% for 0.75 µm. 90 to 99.6% for 3 µm.	>0.5	13 to 250	Reliability may be limited by scaling, fouling, or corrosion. Scrubbing liquor blowdown may require treatment or contain potentially valuable material not directly recoverable.	Scrubbing liquor blowdown and wet slurry	About \$250/m^3/min. Costs vary with particle loading and size distribution, and collection efficiency specified.	High particulate removal efficiency. Capable of treating streams with wide ranges in temperature (no limitation for high temperatures), pressure, and gas composition. High efficiencies require high energy consumption.
Electrostatic Precipitator	A negative electrical charge is imparted to the particulates and they are collected on positively charged plates. Collected material is removed by periodically rapping or vibrating the collection plates.	90 to 99% for 0.1 µm. 80 to 96% for 0.5 µm. 95 to 99% for 1 µm. 99 to 99.9% for 5 µm.	0.24	0.5 to 2.5	Not applicable to combustible or potentially explosive mixtures. Particulates to be collected must have suitable electrical resistivity to facilitate efficient removal. Used in low pressure applications. Limited to gas streams with temperatures below 700 K.	Collected particulates	About \$250 to \$530/m^3/min. Costs vary with particle loading and size distribution, particle resistivity, and collection efficiency specified.	High particulate removal efficiency, especially the sub-micron range. High capital and installation cost. Very low pressure drop. Suitable for high temperature or large volume applications. High electrical consumption. Sensitive to particulate resistivity.

*Capital costs presented in this table are total process costs (TPC) and do not include interest during construction.

Data source: Reference 49

Cyclones are generally employed for the removal of bulk particulates (usually greater than 5 microns in size) and, in many cases, precede other control devices. The unit total process costs (TPC) of cyclones are relatively low - approximately \$212 per m^3/min. Cyclones for particulate control are further discussed in Appendix A-11.

Baghouses have very high particulate removal efficiencies and can lend themselves to applications involving small or intermittent gas flows (although continuous operation is preferred). Baghouses, however, have high pressure drops (in comparison with electrostatic precipitators) and cannot ordinarily handle wet gases, gases containing oily materials, or gases having temperatures in excess of 573°K. The unit total process cost (TPC) for a typical baghouse is about \$300 per m^3/min. Fabric filters for particulate control are further discussed in Appendix A-12.

Venturi scrubbers can generally handle gases having temperatures higher than those which can be handled by fabric filters, can operate at high pressure, can tolerate wet and dry gases, and can be very efficient for the removal of submicron particles (at high pressure drops and hence high operating costs). In contrast to devices in which the particles are collected in dry form, venturi scrubbers generate a wet slurry which is more voluminous and generally more difficult to dispose of. The unit total process costs (TPC) for venturi scrubbers are approximately \$250 per m^3/min. Venturi scrubbers for particulate control are further discussed in Appendix A-14.

Electrostatic precipitators are high efficiency particulate removal devices, have low pressure drops, are capable of handling large volumes of gases, and can tolerate high feed gas temperatures (typically up to 700°K). Electrostatic precipitators, however, are not generally suitable for applications to gases above atmospheric pressure and are not economical for treating small or intermittent gas flows (such as those resulting from material handling dust collection systems). The unit total process costs (TPC) range from \$250 to

\$530 per m^3/min. Electrostatic precipitators for particulate control are further discussed in Appendix A-13.

SO_2 Removal

Several flue gas desulfurization (FGD) processes are commercially available. They are basically of three types: 1) throwaway systems which produce a waste sludge by-product, 2) regenerable systems which produce a usable sulfur by-product and regenerate the sorbent, and 3) dry scrubbing systems. Common examples of throwaway systems are the lime/limestone system, double alkali scrubbing, soda ash scrubbing, fly ash alkalinity scrubbing, and Chiyoda Thoroughbred 121. Typical examples of regenerable systems are the Wellman-Lord and Magnesium Oxide processes. Key features of some of these scrubbing systems are presented in Table 4-30.

The lime/limestone scrubbers are the most commonly employed throwaway systems for electric utility applications. In the lime/limestone process, solid lime or limestone is pulverized and mixed with water to form a scrubber liquor, which contacts the flue gas in an absorption tower where calcium sulfate and calcium sulfite are formed. The resulting slurry is removed from the system and treated, and the sludge is disposed of. Scrubbing solution is recovered and recycled to the absorption tower. Lime/limestone removal efficiencies can approach 90% by carefully balancing the many chemical-reaction parameters involved. Out of the throwaway FGD systems available, lime/limestone offers the least complex system and equipment, the easier pH control, and the cheapest raw materials. Operating experience has indicated that careful attention to system control is important for successful operation of lime/limestone FGD systems.

Of all the commercially available regenerable FGD systems, Wellman-Lord is the most extensively used. A venturi prescrubber often precedes the Wellman-Lord absorber to remove residual particulates from the flue gas, and avoid ash accumulation in the absorber. Sulfur dioxide is absorbed by an alkaline sodium sulfite solution to produce primarily sodium bisulfite. This bisulfite-rich solution is then pumped to a forced-circulation vacuum evaporator where it is

TABLE 4-30. KEY FEATURES OF SO_2 REMOVAL PROCESSES

Process Feature	Lime/Limestone Scrubbing	Double Alkali Scrubbing	Chiyoda Thoroughbred 121	Wellman-Lord	Dry Scrubbing	Fly Ash Alkalinity Scrubbing
Principle	Liquid phase absorption of SO_2 in lime or limestone slurry.	Liquid phase absorption of SO_2 in a sodium hydroxide, sodium sulfite, sodium bisulfite, sodium sulfate, and sodium carbonate solution. Regeneration of the sodium sulfite/bisulfite with lime in a reactor. A dilute mode can be used for concentrations of 200-1500 ppm SO_2 and where less than 25% oxidation of sulfite to sulfate occurs. Concentrated mode can be used for concentrations of 1000-8000 ppm SO_2.	Liquid absorption of SO_2 in a single vessel, where limestone addition and dissolution, air oxidation, and gypsum precipitation occur.	Liquid phase absorption of SO_2 in a sodium bisulfite, sodium sulfite, and sodium carbonate solution. A rich SO_2 is produced by evaporation, which is then processed in a Claus unit to produce elemental sulfur or in a sulfuric acid plant.	Process involves the use of a spray dryer which contacts the flue gas with an aqueous alkaline material and produces a dry product. System involves two stages: 1st stage-spray dryer; 2nd stage-dry particulate collector which removes flue ash and reaction product fron flue gas.	Process involves a two stage venturi-spray tower absorber utilizing the fly ash alkalinity for SO_2 removal. Hydrated dolomitic lime, [$Mg(OH)_2$ and $Ca(OH)_2$] is also used to achieve an outlet SO_2 of 43 ng/J.
Feed Stream Requirements	Particulates can be removed in an ESP or fabric filter to achieve 99+% at lowest energy consumption. Fly ash may be removed in a venturi where the fly ash contains significant alkalinity. A scrubber can be used for both high particulate and SO_2 removal.	Excessive particulates should be removed in an ESP, fabric filter or venturi. O_2 should be less than 7% for concentrated mode.	Particulates and chlorides should be removed from inlet flue gas if byproduct gypsum is to be sold.	Particulates and chlorides must be removed from flue gas.	Inlet SO_2 concentration should not exceed 1000 ppmv.	Venturi is used to remove particulates and a portion of SO_2.
Absorbent	Slaked lime or 200-300 mesh limestone 6-12% slurry circulated.	Sodium hydroxide, sodium sulfite/bisulfite, and a small amount of sodium sulfate.	Limestone slurry.	Concentrated sodium sulfite/bisulfite.	Lime slurry or sodium carbonate solution.	Fly ash alkalinity and hydrated lime (calcium and magnesium hydroxide).
Product/ Waste	Gypsum can be produced with forced oxidation. Calcium sulfate/sulfite can be produced with 50-70% solids.	Filter cake with 60-70% solids, primarily calcium sulfite and calcium sulfate.	Gypsum ($CaSO_4 \cdot 2H_2O$) with less than 20% moisture content.	Concentrated SO_2 purge stream (90% SO_2).	Sodium sulfite-sodium sulfate, calcium sulfite/sulfate.	Sludge consists of fly ash, gypsum ($CaSO_4 \cdot 2H_2O$), $Mg(OH)_2$, small amount of calcium sulfite.
Efficiency	90% removal can be obtained generally for low and high sulfur coals. Higher removals can be obtained with higher L/G and pressure drop, and to some extent scrubber type. 90% and greater can be obtained for low sulfur coals. 95% removal for high sulfur coals when adipic acid is used. Commercially demonstrated in over 30 FGD units.	90-99% removal can be obtained for low and high sulfur coals. Concentrated mode has been demonstrated at Louisville Gas & Electric's 200 MW coal fired boiler. Smaller industrial units (General Motors) have been operated in the dilute mode.	90% SO_2 removal or outlet SO_2 equal to 300 ppmv. Process has been demonstrated at Gulf Power's Scholz station - 20MW prototype.	Process has been demonstrated in a NIPSCO 115MW coal fired boiler. Can remove up to 95% SO_2.	70% SO_2 removal using lime as absorbent. 80% SO_2 removal using sodium absorbent.	85-90% removal of SO_2 commercially demonstrated in Montana Power's Colstrip 1 & 2. 96+% expected in Colstrip units 3 & 4.

(Continued)

TABLE 4-30. (CONTINUED)

Process Feature	Lime/Limestone Scrubbing	Double Alkali Scrubbing	Chiyoda Thoroughbred 121	Wellman-Lord	Dry Scrubbing	Fly Ash Alkalinity Scrubbing
Cost*	Capital - $90/kW to $185/kW.	Capital - $101/kW to $163/kW.	Capital - $160/kW.	Capital - $138/kW to $265/kW.	Capital - $23/kW to $47/kW.	Not available.
Advantages	Lower capital cost and O&M costs. SO_2 and particulates can be removed simultaneously. Relatively simple process. Reliability is 90-95%.	Lower capital and O&M costs. SO_2 can be removed to very high efficiencies (99%). Reliability is 90-95%. Conventional process equipment.	Capital and O&M costs appear to be competitive but data limited to prototype experience. Potential saleable gypsum by-product.	Commercially demonstrated. Process produces saleable product sulfur with a Claus unit or sulfuric acid. Lower potential for scaling than calcium system.	Lower projected operating and capital costs. Dry product.	Commercially demonstrated in 300MW units. Sludge contains little calcium sulfite which improves dewatering and therefore reduces settled water content. Less potential for scaling.
Disadvantages	Process produces approximately 2 times (by wt) sludge as ash collected. Sludge can be thixotropic. Sludge quantities can be reduced by forced oxidation.	Process produces 1.5 times (by wt) filter cake as collected ash. 316SS material of construction may be required to prevent corrosion and pitting.	Process has not been demonstrated commercially in a 100MW or larger unit.	High utility costs (steam) compared to other systems. Special metallurgy may be required. System required to process SO_2 to sulfur or sulfuric acid.	Process will not be commercially demonstrated until late 1982. Product disposal could be a problem when sodium salts are used as absorbent.	Process is generally applied to coal fired boilers which burn high alkalinity coal.

Data source: Reference 45.

* Capital costs included in this table are total process costs (TPC) and do not include interest during construction.

indirectly heated by steam to convert the bisulfite to sulfite and gaseous SO_2. A portion of the sodium sulfite is also converted irreversibly to sodium sulfate and thiosulfate which must be purged from the system, requiring makeup of NaOH or $NaCO_3$. The Wellman-Lord process can achieve over 90% SO_2 removal.

Dry scrubbing experience to date has been limited, although systems that have been operated show much promise, especially for low- and medium-sulfur coals. The spray drying process is the only dry scrubbing process currently being offered commercially. In this process the absorbent solution, usually either lime or soda ash, is atomized and sprayed into the incoming flue-gas stream to increase the liquid/gas interface and promote mass transfer between the SO_2 and the slurry droplets. Simultaneously, the thermal energy of the gas stream evaporates the water in the atomized droplets, to produce a dry, powdered mixture of sulfite/sulfate and unreacted reagent. When used in combination with fabric filters these systems have performed extremely well. The fabric filter collects the particulates and also recovers some of the expensive reagent which is reused. In addition, unused reagent that cakes on the fabric is available to react with more SO_2 as flue gas passes through it.

One type of FGD process not included in Table 4-30 is the sodium scrubbing process. The sodium scrubbing process is capable of achieving high SO_2 removal efficiencies over a wide range of inlet SO_2 concentrations. The process consumes a premium chemical such as Na_2CO_3 and produces a soluble waste salt containing Na_2SO_3 and $NaHSO_3$, which is normally discharged to a lined evaporation pond for drying. Compared with other FGD processes, the sodium scrubbing process is lower in capital investment cost and higher in operating cost because of the simplicity in scrubber design and the expense of sodium chemicals. Based on these considerations, the sodium scrubbing process might be suited for controlling SO_2 emissions from intermittent gaseous waste streams. The capital investment cost (TCI) for sodium scrubbing on a unit capacity basis is approximately \$415 per kmol/hr feed gas, based on a system that removes 90% of the SO_2 contained in the flue gas from a 211 GJ (200 MM Btu) per hr industrial boiler burning 3.5% sulfur coal (45).

FGD costs for boilers in synfuel plants will depend upon the amount of sulfur emissions control required. This may vary depending upon the amount of sulfur in the fuel. FGD cost data has been developed by the EPA for electric utility steam generating units ranging in size from 25 MWe to 1000 MWe. The cost variations are principally governed by: 1) size of the boiler, 2) coal used, 3) averaging time over which the plant must meet SO_2 limitations, and 4) the level of control maintained. Capital investment and annual operating costs for lime/limestone and Wellman-Lord FGD systems are listed in Table 4-30. These costs are for a 500 MWe unit boiler burning 3.5% sulfur bituminous coal capable of achieving 90% removal.

FGD systems are further discussed in Appendix A-20.

4.2.2.1 Flue Gases from Steam and Power Generation Systems (Streams 701a and 707a)

Fuel requirements for the steam generation systems are met by a combination of treated low-Btu fuel gas and coal in current Exxon designs. The option of burning untreated high-sulfur fuel gas in the steam boilers -- and using an FGD system to remove the resulting SO_2 in the flue gas -- is not considered in the current Exxon design, and is not addressed in this manual. For the base case design, total energy input to the steam boiler is 2.32 TJ (2,199 MM Btu) per hour, 63.0% of which is provided by burning low-Btu fuel gas (125 Btu/SCF, 83 ppmv sulfur) and 37.0% provided by burning coal (3.7% sulfur, 7.6% ash) at 35.3 Mg/hr. For the MFS case design, total energy input to the steam boiler is 3.95 TJ (3,740 MM Btu) per hour, 42.8% of which is provided by burning low-Btu fuel gas and 57.2% provided by burning coal at 92.8 Mg/hr. Flue gases from the steam boiler were estimated to be generated at a rate of 42,560 kmol/hr and 69,150 kmol/hr for the base case and MFS case designs, respectively.

Energy for the power generation system, on the other hand, is derived from the combustion of coal alone. For the base case design, the power requirement of 223 MW is met by burning coal at a rate of 95.8 Mg/hr (2.33 TJ/hr). For the MFS case design, the power requirement of 160 MW is met by burning coal at 68.7 Mg/hr (1.67 TJ/hr). Compared with the steam boiler, flue gases from the power boiler are generated at lower rates of 36,450 kmol/hr and 26,110 kmol/hr for the two EDS designs. However, emission concentrations of NO_x, particulate matter, and SO_2 are all significantly higher in the power boiler flue gas because of the exclusive use of coal as combustion fuel.

In this section, pollution control alternatives applicable to the flue gases from the steam and power boilers are examined. The control approach involves all three functions described previously; NO_x control, particulate control, and SO_2 removal. The control of these pollutants from the steam and power boilers in an EDS commercial plant is expected to present no unique problems over those encountered in the electric utility and other industries which use boilers fired by coal and fuel gas.

4.2.2.1.1 Control Function 1 - NO_x Control

As discussed earlier, NO_x control in boilers is achieved through both the design and operation of the combustion units to minimize its formation. For new pulverized coal-fired boilers and gas—fired boilers this is achieved through the application of a variety of combustion modification techniques (see Table 4-27) including low NO_x burners. Boiler manufacturers have regularly incorporated combustion modifications as standard equipment on new boilers since the early 1970's. Accordingly, it is considered in this manual that the cost of any new coal-fired or gas-fired boiler installed in an EDS plant will inherently incorporate the cost of any NO_x combustion modifications that have been incorporated into the boiler. No attempt is made here to separate out the portion of the total cost which is attributable to NO_x control. For the purposes of this manual, it is assumed that a well-designed coal-fired boiler with suitable combustion modifications would be able to meet the current New Source Performance Standard (NSPS) that applies to electric utility coal-fired boilers, 260 ng/J. The NO_x emission factor for low-Btu gas combustion is 52-99 ng/J (Table 3-64), approximately the same as the 86 ng/J NSPS standard for utility boilers burning gaseous fuel. It is assumed that a well-designed gas-fired boiler would be able to operate in the 52-99 ng/J range. For the steam boiler, these emission factors translate into total NO_x emissions of 297-366 kg/hr in the base case, and 670-750 kg/hr in the MFS case. For the coal-fired power boiler, NO_x emissions for these two EDS designs would be 601 kg/hr and 431 kg/hr, respectively.*

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* Note: The use of the electricity utility NSPS in estimating NO_x emissions from the EDS boilers is done here for illustrative purposes only, and is not intended to suggest that EDS boilers would necessarily be required to meet that standard.

4.2.2.1.2 Control Function 2 - Particulate Control

Particulate loading in boiler flue gases is a function of many variables such as the type(s) of fuel being fired, the ash content of the fuel and boiler design. In pulverized coal-fired boilers, typically 60 to 90% of the ash in the feed coal is emitted as fly ash, the remainder appearing as bottom ash. For the steam boiler, uncontrolled emissions of particulate matter were estimated to be 2,190 kg/hr and 5,740 kg/hr for the base case and MFS case designs, respectively. Uncontrolled emissions of particulate matter from the power boiler for these two EDS designs would be 5,930 kg/hr and 4,250 kg/hr.

Electrostatic precipitators (ESPs) or baghouses are commonly used for control of particulate emissions from coal-fired boilers. The use of these technologies has served to meet most state, local or federal performance standards as well as any operational requirements for pollution control equipment employed downstream of these units. In utility applications ESPs are more widely used than baghouses. Only the application of ESPs will be discussed here.

4.2.2.1.2.1 Control Technique 1 - Electrostatic Precipitators

Costs for ESPs are based upon the ESP effluent particulate loading desired and the flue gas flow rate. Typically, to achieve an ESP outlet loading of 13 ng/J, ESP capital investment costs (TCI) are approximately $88.8/kw ($545 per kmol/hr) for Illinois No. 6 coal. For an outlet loading of 43 ng/J, capital investment costs (TCI) decrease to $70.0/kw ($428 per kmol/hr). These costs were derived from the installed equipment costs (IEC) presented in Figure A 13-5 of Appendix A-13 for eastern high sulfur coal, by applying the methodology outlined in Table 4-1 and adjusting to 1980. In achieving either one of these emission levels, over 99% of the particulate matter will be collected by the ESP for the power boiler. Collection efficiencies of the ESP for the steam boiler at these two emission levels would be slightly lower because particulates are generated at a lower rate from the combustion of a combination of coal and fuel gas. Controlled emissions of particulate matter from the low-Btu gas- and

and coal-fired steam boiler (2.32 TJ/hr for the base case, 3.95 TJ/hr for the MFS case) at these two emission levels (13 ng/J and 43 ng/J) were estimated to be 30-100 kg/hr and 51-170 kg/hr for the base case and MFS case designs, respectively. Similarly, controlled emissions of particulate matter from the coal-fired power boiler (2.33 TJ/hr for the base case, 1.67 TJ/hr for the MFS case) at the same two emission levels would be 30-101 kg/hr and 22-72 kg/hr for the two EDS designs.

Estimated capital investment costs and annualized costs for control of particulate matter from steam and power boilers are presented in Table 4-31. On annualized basis, the cost for particulate control with ESP would be \$11 - \$14 per Mmol of flue gas.

Secondary Waste Stream

The fly ash collected is a secondary waste stream generated by the ESP. For the steam boiler, the fly ash collected would amount to 16,740 Mg/yr and 44,420 Mg/yr for the base case and MFS case designs, respectively. For the power boiler, the fly ash collected would amount to 46,240 Mg/yr and 33,120 Mg/yr for these two EDS designs. Concentrations of trace elements present in the fly ash are similar to those found in the coal feed (Table 2-2), although the levels of some of the more volatile elements (or elements associated with volatile compounds) will be enriched (36). Disposal of this secondary waste stream is addressed in Section 4.4.2.6.

4.2.2.1.3 Control Function 3 - SO_2 Removal

As discussed previously, a number of competitive FGD processes are capable of achieving similar reductions. The limestone and Wellman-Lord FGD processes are two of the most widely used SO_2 control processes in the industry today. The costs for these systems vary greatly depending upon the boiler size, coal sulfur, and the degree of SO_2 removal desired. The limestone and the Wellman-Lord FGD processes are discussed here.

TABLE 4-31. COSTS OF PARTICULATE CONTROL WITH ELECTROSTATIC PRECIPITATORS FOR FLUE GASES FROM STEAM AND POWER GENERATION SYSTEMS (STREAMS 701a and 707a)

Cost Element	Base Case		MFS Case	
	Total Capital Investment[*]	Total Annualized Cost[†]	Total Capital Investment[*]	Total Annualized Cost[†]
Steam Boiler (701a)				
Total Cost, $ million	18.2-23.2[‡]	3.6-4.6[‡]	29.6-37.7[‡]	5.9-7.5[‡]
% of Base Plant	0.57-0.73	0.38-0.48	1.02-1.31	0.65-0.83
Unit Cost	$428-545 per kmol/hr flue gas	$11-14 per Mmol/flue gas	$428-545 per kmol/hr flue gas	$11-14 per Mmol flue gas
Power Boiler (707a)				
Total Cost, $ million	15.6-19.8	3.1-4.0	11.2-14.2	2.2-2.8
% Base Plant	0.49-0.62	0.32-0.41	0.39-0.49	0.25-0.31
Unit Cost	$428-545 per kmol/hr flue gas	$11-14 per Mmol flue gas	$428-545 per kmol/hr flue gas	$11-14 per Mmol flue gas

[*] The capital investment costs were obtained using the calculated flue gas flow rates and the unit cost of $70.0-$88.8/kw ($428-545 per kmol/hr flue gas) derived from Figure A 13-5 in Appendix A-13 and using the methodology outlined in Table 4-1.

[†] The annualized costs were derived using the methodology described in Section 4.1.3 and assumptions indicated in Appendix A-13 (Section 6).

[‡] The lower cost corresponds to an emission level of 43 ng/J, and the higher cost corresponds to an emission level of 13 ng/J.

Air Source Type 2
Flue Gases from Steam and
Power Generation System
Control Function 3 -
SO_2 Removal

For the steam boiler, uncontrolled SO_2 emissions were estimated to be 2,520 kg/hr (1095 ng/J) and 6,530 kg/hr (1668 ng/J) for the base case and MFS case designs, respectively. Uncontrolled SO_2 emissions from the power boiler for these two EDS designs would be 6,660 kg/hr (2878 ng/J) and 4,780 kg/hr (2878 ng/J).

4.2.2.1.3.1 Control Technique 1 - Limestone FGD Process

Two different SO_2 emission levels are considered with limestone scrubbing. For the steam boiler, the SO_2 emission level is assumed to be 260 ng/J for the combined gas and coal feed. The 260 ng/J emission level is the NSPS for coal-fired electric utility steam generating units, when uncontrolled SO_2 emissions are in the 867-2,600 ng/J range. This emission level would result in SO_2 emissions of 599 kg/hr and 1,018 kg/hr for the base case and MFS case designs, respectively. For the power boiler, it is assumed that 90% SO_2 removal would be needed. The 90% SO_2 removal requirement is the NSPS for coal-fired electric utility steam generating units, when uncontrolled SO_2 emissions are in the 2,600-5,200 ng/J range. This would result in SO_2 emissions of 666 kg/hr and 478 kg/hr for the two EDS designs. [Note: These emission levels are used here for illustrative purposes only; its use should not be construed to suggest that EDS boilers would necessarily have to meet NSPS].

Estimated capital investment costs and annualized costs for SO_2 removal from steam and power boilers with limestone scrubbing are presented in Table 4-32. On annualized basis, the cost for SO_2 removal with limestone scrubbing would be $52-83 per Mmol of flue gas.

Secondary Waste Stream

The FGD sludge generated from limestone scrubbing typically contains 18.4 wt % $CaSO_4 \cdot 2H_2O$, 45.5 wt % $CaSO_3 \cdot 1/2\ H_2O$, 0.3 wt % fly ash, and 35.8 wt % limestone on dry basis. For the steam boiler, FGD sludge (dry basis) generated would amount to 50,720 Mg/yr and 147,590 Mg/yr for the base case and MFS case

TABLE 4-32. COSTS OF SO_2 REMOVAL WITH LIMESTONE SCRUBBING FOR FLUE GASES FROM STEAM AND POWER GENERATION SYSTEMS (STREAMS 701a and 707a)

Cost Element	Base Case Total Capital Investment*	 Total Annualized Cost †	MFS Case Total Capital Investment*	 Total Annualized Cost †
Steam Boiler (701a)				
Total Cost, $ million	50.2	20.1	70.7	28.3
% of Base Plant	1.57	2.09	2.45	3.13
Unit Cost	$1,179 per kmol/hr flue gas	$60 per Mmol flue gas	$1,023 per kmol/hr flue gas	$52 per Mmol flue gas
Power Boiler (707a)				
Total Cost, $ million	52.0	20.8	42.6	17.1
% of Base Plant	1.63	2.16	1.48	1.88
Unit Cost	$1,427 per kmol/hr flue gas	$72 per Mmol flue gas	$1,632 per kmol/hr flue gas	$83 per Mmol flue gas

* The capital investment costs were obtained using the calculated flue gas flow rates and the unit costs provided in Appendix A-20 (Table A 20-2) for 90% SO_2 removal, and by adjustments to scale down from 500 MW unit in Table A 20-2 based upon the six-tenths rule. For convenience, costs for FGD systems providing less than 90% SO_2 removal (i.e., flue gas from the steam boiler) were estimated by assuming bypassing part of the flue gas and treating the remainder for 90% SO_2 removal so as to achieve the required SO_2 removal.

† The annualized costs were derived from the methodology described in Section 4.1.3, and by using annual operating costs provided in Appendix A-20.

Note: Cost estimates assume one limestone scrubber for steam boiler, and second, separate scrubber for power boiler.

designs, respectively. For the power boiler, FGD sludge (dry basis) generated would amount to 162,450 Mg/yr and 116,500 Mg/yr for these two EDS designs. Disposal of this secondary waste stream is addressed in Section 4.4.2.7.

4.2.2.1.3.2 Control Technique 2 - Wellman-Lord FGD Process

The same SO_2 emission levels are considered here as in the limestone scrubbing case. Estimated capital investment costs and annualized costs for SO_2 removal from steam and power generation with the Wellman-Lord FGD process are presented in Table 4-33. On annualized basis, the cost for SO_2 removal with Wellman-Lord would be \$56 - \$78 per Mmol of flue gas.

Secondary Waste Stream

Sulfur is recovered in the form of elemental sulfur. However, there is a secondary waste stream from the Wellman-Lord process as the thiosulfate/sulfate by-product purge. This by-product purge typically contains 41.5 wt % Na_2SO_3, 18.1 wt % $Na_2S_2O_5$, 10.0 wt % Na_2SO_4, 1.1 wt % $Na_2S_2O_3$, and 29.2 wt % H_2O. For the steam boiler, sulfur would be recovered at a rate of 921 kg/hr and 2,638 kg/hr for the base case and MFS case designs, whereas the by-product purge would be generated at a rate of 209 kg/hr and 598 kg/hr for these two EDS designs. For the power boiler, sulfur would be recovered at a rate of 2,869 kg/hr and 2,057 kg/hr for the base case and MFS case designs, and the by-product purge would be generated at a rate of 650 kg/hr and 466 kg/hr for the two EDS designs. Disposal of this secondary waste stream is addressed in Section 4.4.2.7.

TABLE 4-33. COSTS OF SO_2 REMOVAL WITH THE WELLMAN-LORD FGD PROCESS FOR FLUE GASES FROM STEAM AND POWER GENERATION SYSTEMS (STREAMS 701a and 707a)

Cost Element	Base Case		MFS Case	
	Total Capital Investment*	Total Annualized Cost †	Total Capital Investment*	Total Annualized Cost †
Steam Boiler (701a)				
Total Cost, $ million	48.6	18.9	68.4	31.8
% of Base Plant	1.52	1.96	2.37	3.51
Unit Cost	$1,141 per kmol/hr flue gas	$56 per Mmol flue gas	$990 per kmol/hr flue gas	$58 per Mmol flue gas
Power Boiler (707a)				
Total Cost, $ million	50.3	19.6	41.2	16.1
% of Base Plant	1.58	2.03	1.43	1.77
Unit Cost	$1,380 per kmol/hr flue gas	$68 per Mmol flue gas	$1,579 per kmol/hr flue gas	$78 per Mmol flue gas

* See footnotes for Table 4-32.

Air Source Type 2
Flue Gas from Process Heaters

4.2.2.2 Flue Gases from Process Heaters (Streams 107, 161, 203, 434)

There are three major flue gas streams from process heaters for both the base case and the MFS case designs:

- Stream 107 - flue gas from liquefaction slurry preheate furnace
- Stream 161 - flue gas from partial oxidation feed vacuum fractionator preheat furnace (MFS case only)
- Stream 203 - flue gas from solvent hydrogenation fuel preheat furnaces
- Stream 434 - flue gas from hydrogen plant reformer furnaces (base case only).

These flue gas streams would be expected to be generated by the combustion of the low Btu fuel gas from Flexicoking. Uncontrolled emissions of particulate matter were estimated to be 2.1 - 6.4 ng/J, compared to an NSPS of 13 ng/J for electric utility steam generating units; the low emission results because the fuel gas is expected to have a low particulate loading. Emissions of NO_x from the process heaters were estimated to be 52-99 ng/J, compared to an NSPS of 86 ng/J for electric utility steam generating units. Thus, there does not appear to be further need for control of particulate matter and NO_x.

Sulfur emissions from these furnaces could be controlled by one of two alternative approaches: removal of reduced sulfur from the Flexicoking sour fuel gas prior to combustion, or removal of SO_2 from the furnace flue gas after combustion. Treatment of the Flexicoking sour fuel gas for removal of reduced sulfur compounds by the Stretford process is discussed in Section 4.2.1.3. The alternative of installing flue gas desulfurization systems on the furnace flue gas appears to be less attractive from both technical and economical considerations. The capital investment cost and the annualized cost of 90% SO_2 removal with limestone scrubbing would be approximately 3.5 and 4.6 times the corresponding costs for sulfur removal with the Stretford process. Further, over 97% of the reduced sulfur compounds in the Flexicoking sour fuel gas would be removed by the Stretford process. If treated by Stretford, the fuel gas would contain 83 ppmv sulfur, which corresponds to SO_2 emissions of 48 ng/J. By comparison, SO_2 emissions after 90% removal with FGD systems would be approximately 250 ng/J.

4.2.2.3 Regeneration/Decommissioning Off-Gas from High and Low Temperature Shift Catalysts (Streams 449/450)

Off-gas from regeneration/decommissioning of high and low temperature sour shift catalysts with air and steam is generated only from the MFS case design. This is an intermittent waste stream estimated to be generated at a rate of 22,255 kmol/hr. Regeneration would be expected once a year, and up to 72 hours in duration each time. The waste stream would consist primarily of steam with 6% N_2, 1% SO_2 and 0.5% CO_2.

Since the shift gases are provided with interstage and effluent cooling and condensation, it is entirely reasonable to expect operation of the cooling and quench systems during catalyst regeneration to condense moisture from the off-gas. Removal of most of the moisture would be desirable for purposes of volume reduction, prevention of condensation in piping to control devices, and to avoid the other problems which arise in controlling gas streams with high moisture levels. At 316°K (110°F), the off-gas would be reduced to a flow rate of 1877 kmol/hr. This gas stream would contain 8.7% H_2O, 13.0% SO_2, 72.4% N_2, and 5.9% CO_2.

For this off-gas stream, the only control function needed would be SO_2 removal.

4.2.2.3.1 Control Function 1 - SO_2 Removal

Sulfur dioxide removal in the off-gas after bulk moisture removal can be accomplished by a number of FGD processes that are commercially available. Alternatively, this off-gas could also be sent to the Claus plant, so that the SO_2 in the shift catalyst regeneration off-gas would displace SO_2 produced by air oxidation of H_2S in the combined acid gas. Since SO_2 from the shift catalyst regeneration off-gas is about 245 kmol/hr, while H_2S in the combined acid gas feed to the Claus plant is 921 kmol/hr, there is sufficient capacity in the Claus to allow substitution of the SO_2 in the off-gas for SO_2 produced by air oxidation. For the MFS case design, the introduction of the shift catalyst regeneration off-gas into the Claus plant would only increase the flow rate of the Claus tail gas by approximately 2%. Thus, control of the shift catalyst regeneration off-gas by the Claus process is a potential option if fluctuations in sulfur loading caused by Streams 449/450 and if control of proper air feed to the Claus H_2S oxidation step can be accommodated in the Claus design and operation.

It should be commented that there are currently a number of installations using sour shift catalysts in petroleum and related applications. As far as is known, control of catalyst regeneration off-gases is not practiced at any of the facilities. Catalyst users indicate that, due to the short duration and high variability of SO_2 containing off-gases, control would be difficult. Regeneration often occurs when facilities are shut down for general maintenance, so that use of FGD or Claus units onsite at such time presents difficulties. Even if regeneration of catalyst units were to be scheduled during times in which FGD or Claus units were online, integration of the highly variable regeneration gases into the FGD or Claus units could cause difficulties in overall system control. For these reasons, only the sodium throwaway FGD process dedicated to SO_2 control for intermittent waste gas streams is discussed here.

4.2.2.3.1.1 Control Technique 1 - Sodium Throwaway Process

As discussed previously, the sodium throwaway process is lower in capital investment cost and higher in operating cost compared to other FGD systems. For the control of intermittent waste gas streams, however, the high cost of sodium carbonate employed for scrubbing is a lesser concern. Sodium throwaway processes can be designed to remove 90% of the SO_2 present in the inlet stream. Thus, sodium throwaway processes can reduce the SO_2 concentrations in the shift catalyst regeneration off-gas to 1.3% after moisture removal. This is equivalent to an SO_2 concentration of 0.1% in the off-gas prior to moisture removal. Higher SO_2 removal by sodium throwaway processes can be readily achieved because of the high reactivity of sodium chemicals with SO_2.

Estimated capital investment cost and annualized cost for the sodium throwaway process are presented in Table 4-34. As noted in Section 4.2.1.5, any sodium throwaway FGD system sized for the control of SO_2 emissions from the transient waste gases would have much greater capacity and could also be used to control SO_2 emissions from the sour shift catalyst regeneration off-gas. Treatment of Streams 449/450 in any scrubber installed for the transient waste gases would eliminate the costs shown in Table 4-34, except for a small fraction attributable to cost of sodium carbonate. Almost all the costs for treatment of Streams 449/450 would then be included in the costs shown in Table 4-17.

TABLE 4-34. COSTS OF SO_2 REMOVAL WITH SODIUM THROWAWAY PROCESS FOR SOUR SHIFT CATALYST REGENERATION/ DECOMMISSIONING OFFGAS (STREAMS 449/450)

Cost Element	MFS Case Total Capital Investment*	 Total Annualized Cost†
Total Cost, $ million	0.99	0.86
% of Base Plant	0.038	0.095
Unit Cost	$45 per kmol/hr waste gas	$0.54 per kmol waste gas

* The capital investment costs were obtained using che calculated combustion product flow rate and cost data provided in Reference 45, which indicate a unit capital investment cost (TCI) of $530 per kmol/hr combustion product gas for units that handle approximately 2000 kmol/hr gas flow.

† The annualized costs were derived using the methodology described in Section 4.1.3. Sodium carbonate consumption was assumed to be 1.2 mol per mol SO_2 removed. Cost of sodium carbonate was assumed to be $264/Mg.

4.2.3 Source Type 3 - Organic and CO Containing Waste Gases

There are four waste streams for the base case design and one waste stream for the MFS case design under Source Type 3. These waste streams are:

- Stream 426 - vent gas from CO_2 removal by Catacarb process (base case only)
- Stream 438 - hydrogen plant deaerator vent
- Stream 446 - regeneration/decommissioning off-gas from reformer catalyst (base case only)
- Stream 448 - decommissioning off-gas from methanation catalyst (base case only).

The vent gas from CO_2 removal are expected to contain organics and CO at the hundreds of ppmv level. The other three waste streams are expected to contain only traces of hydrocarbons and/or CO, but the exact concentration levels are not known.

Because waste streams under Source Type 3 are free of sulfur compounds, the only control function needed would be incineration for removal of organics and CO. Incineration processes applicable to the control of these waste streams include thermal incineration, catalytic incineration, and flaring. Key features of these incineration processes are discussed in Section 4.2.1. Details regarding these processes are included in Appendices A-15 through A-17.

In an EDS commercial plant, waste gases discussed in this section might be combined with other waste gases for incineration in a common incinerator. However, because of the limited characterization and generation rate data generally available for these waste gases, cost estimates in this section are presented in terms of dedicated incineration units. Therefore, these control costs represent upper limit costs for control of the waste gases. Costs experienced in practice could be lower than these costs if a single, common incinerator were employed which would be designed to handle any one of the streams individually, or any combination of the streams which might reasonably be expected to be generated concurrently.

4.2.3.1 Vent Gas from CO_2 Removal (Stream 426)

The vent gas from CO_2 removal (by Catacarb or alternative AGR system) is generated at a rate of 6100 kmol/hr, and applies only to the base case design. This vent gas will consist primarily of carbon dioxide (78.4%) and steam (19.6%), along with 1.9% hydrogen, 370 ppmv methane and 160 ppmv carbon monoxide. Direct discharge of this vent gas to the atmosphere has been considered to be one of the options. In the event that further reduction of hydrocarbon and CO emissions would be needed, incineration is a control function that may be considered.

4.2.3.1.1 Control Function 1 - Incineration

Thermal and catalytic incineration are applicable techniques to control organic and CO emissions from the CO_2 vent gas. These control techniques are discussed in this section.

4.2.3.1.1.1 Control Technique 1 - Thermal Incineration

The CO_2 vent gas was estimated to contain 2.2 kmol/hr of methane and 0.96 kmol/hr of CO. Thermal incineration of organic-containing waste gases and liquids can typically result in flue gas concentrations ranging from 5 to 20 ppmv total hydrocarbons (expressed as CH_4) and 10 to 100 ppmv CO. Assuming incineration of the CO_2 vent gas at 1100°K (1520°F) with 5,330 kmol/hr (0.59 TJ/hr) of a low-Btu fuel gas and 50% excess air, the generation rate of the flue gas stream would be 18,510 kmol/hr. At concentration levels of 20 ppmv total hydrocarbons and 50 ppmv CO, this flue gas would result in emissions of 0.37 kmol/hr of total hydrocarbons (expressed as CH_4) and 0.93 kmol/hr of CO. Thus, thermal incineration would reduce emissions of total hydrocarbons by 83%. There would be no net reduction in CO emissions because of the additional CO contributions from combustion of the low-Btu fuel gas.

Estimated capital investment and annualized costs for thermal incineration (with waste heat recovery) of the CO_2 vent gas are presented in Table 4-35. Because this stream is generated on a continuous basis, and utilizes a

TABLE 4-35. COSTS OF ORGANICS AND CO CONTROL BY THERMAL INCINERATION WITH WASTE HEAT RECOVERY FOR CO_2 VENT GAS (STREAM 426)

Cost Element	Base Case Total Capital Investment*	 Total Annualized Cost†
Total Cost, $ million	3.2	17.1
% of Base Plant	0.10	1.78
Unit Cost	$516 per kmol/hr waste gas	$356 per Mmol waste gas

* The total capital investment cost (TCI) was obtained using the calculated combustion product flow rate and the installed equipment cost curve (Curve B) presented in Figure A 16-6 of Appendix A-16. This capital investment cost corresponds to a unit cost of $170 per kmol/hr combustion product gas.

† The annualized cost was derived using the methodology described in Section 4.1.3. The annual cost for fuel gas is $20.5 million, and the annual steam credit received is $4.0 million.

substantial quantity of low-Btu gas, the annualized cost for thermal incineration is relatively high and represents 1.8% of the total annualized cost for the uncontrolled EDS base plant.

4.2.3.1.1.2 Control Technique 2 - Catalytic Incineration

The efficiency of catalytic incinerators in destroying organics and CO is strongly dependent on the type of catalyst used and the temperature of the catalyst bed. At a temperature of 700°K (800°F), the destruction efficiencies for most organics and for CO are over 97% and approximately 99%, respectively (Appendix A-17). Effluent concentrations for total hydrocarbons are typically 1 to 20 ppmv. Catalytic incineration of the CO_2 vent gas at 700°K (800°F), with 1,230 kmol/hr (0.28 TJ/hr) of a low-Btu fuel gas and 50% excess air, would generate a flue gas stream at a rate of 11,990 kmol/hr (170,000 SCFM). At concentration levels of 20 ppmv total hydrocarbons and 50 ppmv CO, this flue gas would result in emissions of 0.24 kmol/hr of total hydrocarbons (expressed as CH_4) and 0.60 kmol/hr of CO. This is equivalent to approximately 89% net reduction in the emission of hydrocarbons. Net reduction in CO emissions is not anticipated because of additional CO contributions from combustion of the low-Btu fuel gas, and because actual CO emissions from any incineration device could fluctuate over a wide range (e.g., 10-100 ppmv) during even a short time period.

Estimated capital investment and annualized costs for catalytic incineration of the CO_2 vent gas are presented in Table 4-36. The capital investment cost for catalytic incineration would be slightly higher than that for thermal incineration. The total annualized cost for catalytic incineration, however, would be 39% lower than that for thermal incineration.

TABLE 4-36. COSTS OF ORGANICS AND CO CONTROL BY CATALYTIC INCINERATION FOR CO_2 VENT GAS (STREAM 426)

Cost Element	Base Case: Total Capital Investment*	Base Case: Total Annualized Cost †
Total Cost, $ million	3.6	10.4
% of Base Plant	0.11	1.09
Unit Cost	$595 per kmol/hr waste gas	$217 per Mmol waste gas

* The capital investment cost was obtained using the calculated combustion product flow rate, the equipment cost curve for fixed-bed catalytic incineration presented in Appendix A-17 (Figure A 17-3), and the methodology outlined in Table 4-1.

† The annualized cost was derived using the methodology described in Section 4.1.3. The annual cost for fuel gas amounts to $9.7 million.

4.2.3.2 Hydrogen Plant Deaerator Vent (Stream 438)

The deaerator vent gas from the hydrogen plant is generated at a rate of 6.5 kmol/hr for the base case design and 6.2 kmol/hr for the MFS case design. This vent gas would consist primarily of steam with a small quantity of carbon dioxide and traces of light hydrocarbons. Direct discharge of this vent gas to the atmosphere is one of the options. However, this waste stream and other vent gases are usually not well characterized, and incineration is a control function that might be considered if organic and CO emissions are of concern.

4.2.3.2.1 Control Function 1 - Incineration

Thermal incineration, catalytic incineration, and flaring are applicable techniques to control organic and CO emissions from vent gases. Only thermal incineration is discussed in this section, because this is a small volume waste gas stream.

4.2.3.2.1.1 Control Technique 1 - Thermal Incineration

As discussed previously, thermal incineration of waste gases would typically result in flue gas concentrations of 5-20 ppmv total hydrocarbons (as CH_4) and 10-100 ppmv CO. Based on energy balances, thermal incineration of the hydrogen plant deaerator vent would require approximately 0.72 mol of low-Btu fuel gas (125 Btu/SCF) per mol of waste gas, assuming incineration at 1100°K (1520°F) with 50% excess air. The estimated unit capital investment (TCI) cost, derived using the calculated combustion product flow rate and the installed equipment cost (IEC) curve for thermal incineration presented in Appendix A-16, (Curve A, Figure A 16-6), would be about $2600 per kmol/hr of waste gas. The estimated unit annualized cost, derived using the cost methodology described in Section 4.1.3 and the fuel gas requirement, would be $490 per Mmol waste gas. Thus, the capital investment and annualized costs for thermal incineration of the hydrogen plant deaerator vent would be approximately $16,900 and $25,000, respectively. These unit costs can also be used to estimate the capital investment and annualized costs for thermal incineraation of other small volume vent gases in dedicated units.

4.2.3.3 Catalyst Regeneration/Decommissioning Off-Gases (Streams 446/448)

There are two off-gas streams from the regeneration and/or decommissioning of catalysts for the base case design. For the reformer catalyst, it has been estimated that regeneration will occur over a period of 24 hours per year, and the off-gas will be generated at an average rate of 6,800 kmol/hr. The off-gas is anticipated to consist primarily of steam and nitrogen with small amounts of CO and particulate matter. Methanation catalysts are not typically regenerated; however, these catalysts are oxidized during decommissioning because they tend to be pyrophoric in the active state. The decommissioning off-gas has been estimated to be generated at an average rate of approximately 5,000 kmol/hr. The average emission frequency of this off-gas is once every four years, with an average emission duration of 24 hours. The characteristics of the decommissioning off-gas are not well known, except that it is expected to consist primarily of steam and nitrogen with small amounts of particulate matter. For either off-gas stream, no data are available on CO or particulate levels. Residual organics may also be present during purging of the unit or initial phases of regeneration/decommissioning.

Control of the catalyst regeneration/decommissioning off-gases by incineration is discussed in this section.

4.2.3.3.1 Control Function 1 - Incineration

Flaring and thermal incineration are applicable techniques to control organic and CO emissions from catalyst regeneration/decommissioning off-gas. These control techniques are discussed in this section. Catalytic incineration is not applicable because the presence of particulate matter in these off-gases may result in catalyst deactivation.

4.2.3.3.1.1 Control Technique 1 - Flaring

Data from recent studies indicate that 96 to >99% of the hydrocarbons in waste gases are typically destroyed by flaring (Appendix A-15). Data on the effectiveness of CO control by flaring are limited; CO levels of hundreds of ppmv have been reported.

Estimated capital investment and annualized costs for flaring of catalyst regeneration/decommissioning off-gases are presented in Table 4-37. These estimates are based on the assumptions that the off-gases would not have significant heating value, and that a low-Btu fuel gas would be employed as supplemental fuel.

4.2.3.3.1.2 Control Technique 2 - Thermal Incineration

Concentrations of CO and hydrocarbons in the incinerated gas would typically range from 10-100 ppmv and 5-20 ppmv, respectively. Although characterization data for the catalyst regeneration/decommissioning off-gases are not available, emissions of CO and hydrocarbons from the incineration of these gases may be estimated by assuming typical flue gas concentrations of 50 ppmv CO and 20 ppmv total hydrocarbons. Incineration of the reformer catalyst regeneration/decommissioning off-gas with a low-Btu fuel gas at 1100°K (1520°F) and 50% excess air would generate 14,850 kmol/hr of flue gas, 0.30 kmol/hr of total hydrocarbons and 0.74 kmol/hr of CO. Similarly, incineration of the methanation catalyst decommissioning off-gas would generate 10,920 kmol/hr of flue gas, 0.22 kmol/hr of total hydrocarbons and 0.55 kmol/hr of CO.

Estimated capital investment and annualized costs for thermal incineration of catalyst regeneration/decommissioning off-gases are presented in Table 4-38. These estimates are also based on the assumption that these off-gases would not have significant heating value.

TABLE 4-37. COSTS OF ORGANICS AND CO CONTROL BY FLARING FOR CATALYST REGENERATION/DECOMMISSIONING OFF-GASES

Cost Element	Reformer Catalyst Regeneration/ Decommissioning Off-Gas		Methanation Catalyst Decommissioning Off-Gas	
	Total Capital Investment*	Total Annualized Cost†	Total Capital Investment*	Total Annualized Cost†
Total Cost, $ million	1.5	0.29	1.1	0.21
% of Base Plant	0.046	0.030	0.038	0.023
Unit Cost	$216 per kmol/hr waste gas	$1.8 per kmol waste gas	$216 per kmol/hr waste gas	$7.1 per kmol waste gas

* The capital investment costs were derived using the combined waste gas and low-Btu fuel gas flow rates and the equipment cost curve for 200 ft high flares presented in Appendix A-15 (Figure A 15-3).

† The annualized costs were derived using the methodology described in Section 4.1.3 and the calculated fuel gas requirements.

TABLE 4-38. COSTS OF ORGANICS AND CO CONTROL BY THERMAL INCINERATION FOR CATALYST REGENERATION/ DECOMMISSIONING OFF-GASES

	Reformer Catalyst Regeneration/ Decommissioning Off-Gas		Methanation Catalyst Decommissioning Off-Gas	
	Total Capital Investment *	Total Annualized Cost †	Total Capital Investment *	Total Annualized Cost †
Total Cost, $ million	2.0	0.38	1.5	0.29
% of Base Plant	0.062	0.040	0.046	0.030
Unit Cost	$291 per kmol/hr waste gas	$2.3 per kmol waste gas	$291 per kmol/hr waste gas	$9.5 per kmol waste gas

* The capital investment costs were obtained using the calculated combustion product flow rates and the installed equipment cost curve presented in Appendix A-16 (Curve A, Figure A 16-6). This capital investment cost corresponds to a unit cost of $133 per kmol/hr combustion product gas.

† The annualized cost was derived using the methodology described in Section 4.1.3 and the calculated fuel gas requirements.

4.2.4 Source Type 4 - Fugitive Dust from Material Storage

There is one uncontrolled waste stream identified under Source Type 4:

Stream 011 - fugitive dust from coal pile

Another waste stream in this category is a secondary waste stream: fugitive dust emissions from solid waste storage piles. Since these waste streams are identical to similar waste streams from conventional sources such as coal-fired power plants and because the control techniques are not stream-specific, a general discussion of applicable control techniques is given rather than a stream-by-stream presentation.

Open or partially enclosed storage piles are often used for bulk materials not affected by precipitation or slight contamination such as coal, sand, gravel, clay, and gypsum. The material may be stored for a short time with a high turnover rate to accommodate surges in daily or weekly rates of sequential processes, or may provide a long-term reserve for emergency supply or to meet cyclical seasonal demands.

Most dust arises from stockpile areas as the material is dumped from the conveyor or chute onto the pile and as material is reclaimed from the pile. During periods of high wind speeds or low moisture, wind erosion of the surface may also cause emissions.

In EDS based commercial plants fugitive dusts are emitted from coal and solid waste storage piles. The techniques used to control these emissions are not unique to liquefaction plants and are widely used in industries that require large scale material storage. During loading operations, fugitive dust emissions could be reduced by minimizing the distance from free drop to the stockpile, or by incorporating devices such as a telescoping chute to essentially enclose the coal stream (or solid waste stream) during the drop. Water sprays could also be directed into the unloading coal or waste and the storage pile. During reclamation, water spraying could again be used for control of fugitive dusts. The most commonly used techniques to control dust emissions from storage piles are water spraying, chemical stabilization, stacked segregation, confinement, capping, and vegetative stabilization. Surface protection methods such as vegetative stabilization, chemical stabilization, capping and stacked segregation are primarily used on reserve storage piles since these piles are subject to minimal disturbances. Active storage piles generally require either water spraying or confinement to control dust emissions. The key features and unit costs for these techniques are listed in Table 4-39. These techniques apply equally to coal storage piles and solid waste storage piles, including methods such as vegetative stabilization, which have been used by the Tennessee Valley Authority on coal piles.

Water spraying is common method of dust suppression. Dust control by water spraying is usually obtained by placing spray nozzles at strategic locations over the stockpile area. The spraying operation is simple in that it only involves the operation of a pump. Water requirements for large volume operations vary from 210 to 250 liters/Mg of material. Such systems are, however, prone to freezeups during winter months. Also, the added moisture can create handling problems during reclamation and subsequent processing. Water spraying is also commonly employed during loading and reclamation operations.

Chemical stabilization to decrease fugitive dust emissions involves the application of wetting or crusting agents. Wetting agents are used to provide better wetting of fines and longer retention of moisture. They also reduce the water surface tension allowing the fines to be wetted with a minimum amount of water. This treatment protects stockpiled material until the added moisture is removed by heat and wind. Some of these agents remain effective for weeks or months without additional rewatering, depending on local conditions. Crusting procedures involve the use of bunker C crude oil, water soluble acrylic polymers, or organic binders. These materials are sprayed on

TABLE 4-39. KEY FEATURES OF STORAGE PILE DUST CONTROL TECHNOLOGIES

Method	Control Principle	Control Effectiveness*	Reliability/ Special Problems	Unit Costs*	Other Pollutants Generated
Water Spray	Spray application of 210-250 l/Mg to reduce dusting	Approximately 50% reduction in losses	1. Piping may require heat tracing when freezing is a concern 2. May increase degradation 3. Frequent re-treatment necessary		none
Chemical	Wetting Agents Modify surface tension properties to improve effectiveness of water sprays	Up to 90% reduction in dust losses‡	1. Piping may require heat tracing when freezing is a concern 2. Can cause corrosion problems in equipment exposed to sprays 3. May increase material degradation 4. Effects are short-term	$.33-.77/Mg†	Volatiles which depend on wetting agent utilized
	Crusting Agents Organic binders combine with particles to form tough crust on surface	Up to 90% reduction in dust losses	1. May increase chances of spontaneous combustion, especially in piles, subject to stockpile and reclaim operations 2. Crust tends to break up during heavy rains	$.55-.22/m^2	Volatiles which depend on crusting agent utilized

(Continued)

TABLE 4-39. (Continued)

Method	Control Principle	Control Effectiveness*	Reliability/ Special Problems	Unit Costs*	Other Pollutants Generated
Stacked Segregation	Coating surface of compacted storage pile with layer of select, medium sized material	No data available	Either deliveries of different sized material must be coordinated or both sizes must be readily available from storage		none
Confinement	Enclosure of active storage pile in a totally enclosed barn or silo with point source dust control equipment on building vents	Up to 99% reduction in losses		a) $110/Mg of stored material b) $1 million to $3 million per silo depending on size	none
Capping	Paving with earth or asphalt or cover with polyethylene	Up to 100%	1. Both coverings may increase chances of spontaneous combustion 2. Polyethylene presents severe handling problems and is also not practical in high wind climates	$.49/m^2 for asphalt $1.96/m^2 for polyethylene	Soil dust from earth covering
Vegetative Stabilization	Covering pile with sod	Approximately 65% reduction over unstabilized pile; 90% if chemical stabilizer is also used	1. Requires frequent watering 2. Handling of sod during reclamation operations is cumbersome and expensive 3. Upper layer of stored material is contaminated with soil	$2.70/m^2	Soil dust from earth covering

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* Cost and control efficiency data obtained from Reference 50.

† Data obtained from Reference 51.

‡ Data obtained from Reference 15.

the surface of the storage pile, coating the top layer of particles with a thin film. This film causes the particles to adhere to one another, forming a tough durable crust which is resistant to wind and rain. As long as the crust remains intact, the storage pile is protected from wind losses.

Another effective means of controlling dust emissions from material storage piles is the stacking of coarse material on the surface of a properly compacted pile. For instance, a 0.152 m layer of fine coal (6.4 mm x 0 mm) on the top and sides of the coal storage pile can be anchored in place by a 0.102 m layer of larger size coal (24 mm x 0 mm) placed on top of the fine coal. The larger size coal has better weathering characteristics compared to the smaller sized coal.

Enclosure of the material storage pile is generally the most effective means of reducing fugitive dust emissions, because it allows the emissions to be captured. However, enclosures can be very expensive, since they have to be designed to withstand wind and snow loads and meet requirements for interior working conditions. An alternative to enclosure of all material is to screen the material prior to storage, sending the oversize material to open storage and the fines to enclosures.

Capping involves the paving of the surface area of the storage pile with asphaltic compounds or earth, or covering the pile with polyethylene tarpaulins. Usually a slurry of wood pulp and asphalt, or road tar is sprayed over the surface of the pile. The covering is usually about 3 mm thick. Polyethylene tarpaulins are also used; however, they are cumbersome to handle when there are high wind speeds or when a large size storage pile is to be capped.

Vegetative stabilization involves planting an appropriate ground cover or shrub over the pile to be stabilized. A soil cap may be required to support vegetation. The efficiency of vegetative cover in reducing wind erosion is dependent on the density and type of vegetation that can be grown. For applications such as stabilizing tailing piles, the use of vegetative stabilization results in a decrease in emissions of approximately 65%. When vegetative stabilization is used in conjunction with a chemical stabilizer, this efficiency increases to approximately 90%.

There are, of course, other fugitive dust emission control techniques that have not been addressed here. One possibility, for example, is to use windscreens (porous windbreaks) to reduce the velocity of the air in the vicinity of the storage pile. Another technique is the optimal profiling and orientation of the storage piles.

4.2.5 Source Type 5 - Fugitive Organic Emissions

There are two types of fugitive organic emissions in an EDS commercial plant:

- Stream 751 - evaporative emissions from product and by-product storage
- Fugitive organic emissions from process equipment such as pump and compressor seals, valves, pipe flanges, pressure relief valves, and drains.

Because of the differences in the nature of these emissions, control techniques for each type are discussed in individual subsections.

Evaporative Emissions from Product and By-Product Storage

Evaporative emissions from storage tanks storing volatile liquids occur because of temperature change, pressure change, or wind induced mechanisms which cause vapor pressure of the stored liquid to vary, resulting in vapor emissions. The minimum accepted standard for storage of volatile liquids is the fixed roof tank. Fixed roof tanks are commonly equipped with a pressure relief valve that allows them to operate at a slight internal pressure or vacuum. The pressure/vacuum valves prevent the escape of vapors only when the changes in temperature, pressure or liquid levels are relatively small. With greater changes in these variables, vapors would be emitted through the pressure/vacuum valves. These vapor emissions are contributed primarily by breathing losses and working losses (21). Breathing loss is the expulsion of vapor from a tank due to vapor expansion and contraction from changes in temperature and barometric pressure. It occurs in the absence of any liquid level change in the tank. Working loss is the combined loss from filling and emptying. In filling, the vapors are expelled from the tank when the pressure inside the tank exceeds the relief pressure, as a result of increase of the liquid level in the tank. In emptying, air is drawn into the tank and becomes saturated with organic and expands, thus exceeding the capacity of the vapor space.

Emissions from fixed roof tanks can be reduced by minimizing diurnal temperature variations (e.g., placing tanks underground), proper setting and maintenance of pressure/vacuum vents, and leak prevention efforts. Significant controls can be effected by: 1) floating a cover on the surface of the stored liquid, 2) using vapor processing units, or 3) by replacing the fixed roof storage tank with a floating roof storage tank.

Floating roof tanks (external and covered) successfully limit hydrocarbon losses by eliminating the ullage into which stored material vaporizes. This is accomplished by floating a rigid deck or roof on the surface of the stored liquid, thus eliminating air space and preventing the formation of organic vapor above the liquid surface. Evaporative losses from floating roof tanks are mainly due to standing storage losses and withdrawal losses (21). Standing storage loss results from wind induced mechanisms as air flows across the top of an external floating roof tank. Withdrawal loss is the vaporization of liquid that clings to the tank wall, and is then exposed to the atmosphere when a floating roof is lowered by withdrawal of liquid. To effectively control emissions, the floating roof employs a seal system to cover the small annular space between the roof and the tank wall. Floating roof tanks are commonly equipped with primary seals. However, evaporative losses from floating roof tanks can be greatly reduced by the installation of secondary seals. The secondary seal is generally of a resilient fabric (e.g., loop seals) or a pliable material such as a treated rubber. Its flexibility allows it to maintain contact in places where the tank shell (walls) might be slightly out of round, as well as in areas where rivet heads project from the shell wall. Upon descent of the roof, these seals wipe down the film left behind by the primary seal. These seals also reduce standing storage evaporative losses since they form a second seal above the vaporized product which has diffused past (or permeated through) the primary seal. Not only do they form a second barrier for the vapor, they also seal this vapor off from the effects of moving air. As a result, secondary seals are effective control devices which, when used on floating roof tanks, can reduce overall emissions by as much as 95% when compared to the use of only primary seals.

Vapor processing units can also be used to control VOC emissions from the pressure/vacuum vents in fixed roof storage tanks. Some of the vapor processing techniques available are carbon absorption, thermal oxidation, refrigeration, compression-refrigeration-absorption, and compression-refrigeration-condensation.

The carbon adsorption vapor recovery unit uses beds of activated carbon to remove VOCs from the air-vapor mixture, that is vented from the overhead space in the tank. These units generally consist of two vertically positioned carbon beds and a carbon regeneration system. Air-vapor mixture enters the base of one of the adsorption columns and the VOC components are adsorbed onto the activated carbon as the gases ascend. Adsorption in one carbon bed occurs for a specific timed cycle before switchover to desorption occurs. The nearly saturated carbon bed is then subjected to vacuum, steam, or thermal regeneration, or a combination of these methods, and the VOCs are stripped from the bed. Vacuum regenerated units recover VOCs by absorption in a gasoline stream which circulates between the control unit and gasoline storage. Some vacuum regenerated systems remain in operation for up to two hours after loading activity ceases, in order to collect any residual vapors in the system and to assure complete regeneration of the carbon beds. The air and any remaining VOCs exiting from the absorber are passed again through the absorbing bed, and exhausted to the atmosphere. Steam regenerated units condense the VOC-water mixture of steam and stripped VOCs and return the separated organic product to storage.

Thermal oxidation units rely upon burning VOC vapors to produce non-polluting combustion products. Vapors driven from the absorbent by increased temperature are piped either to a vapor holder or directly to the oxidizer unit. When a vapor holder is used, operation of the oxidizer begins when the holder reaches a preset level, and ends when the holder is empty. With no vapor holder in the system, the oxidizer is energized by means of pressure in the vapor line, or by an electrical signal produced by manual activation. In some cases propane is injected into the vapor stream to keep the VOC level above the explosive range.

Refrigeration type recovery units remove VOCs from the air-vapor mixture by straight refrigeration at atmospheric pressure. Vapors displaced from storage tanks enter a condenser section where methylene chloride "brine" is pumped through the finned tube sections of the heat exchanger. Brine temperature in this section ranges from 190°K to 210°K. Some units contain a precooler section (glycol and water solution circulating at 274°K) to remove most of the water from the vapors prior to the main condenser. There are no compression stages in this type of unit. The condensed product is collected and pumped to one of the product storage tanks. The cold collection surfaces are periodically defrosted by pumping warm (305°K) trichlorethylene through the condenser. This defrost fluid is kept warm by heat salvaged from the refrigeration equipment. Recovered water passes to a waste storage tank or gasoline-water separator. The defrost cycle takes from 15 to 60 minutes, depending on the amount of ice accumulated on the finned-tubes.

In a compression-refrigeration-absorption (CRA) vapor recovery system, the vapors from the storage tanks are first passed through a saturator which sprays liquid gasoline into the air-vapor gas stream. This ensures that the VOC concentration is above the explosive range. The saturated gas mixture is stored in a vapor holder until, at a preset level, it is released to the control unit. The vapor holder is usually a special tank containing a bladder with variable volume and constant pressure. A product storage tank with a lifter roof can also function in this capacity.

The first stage of CRA processing is a compression-refrigeration cycle in which the vapors are compressed and cooled, and water and heavy VOCs are condensed. The uncondensed vapors move into a packed absorber column where they are contacted by chilled gasoline (277°K) drawn from product storage, and absorbed. The fresh product stream is used first in the saturator, then it passes through an economizing heat exchanger as it enters the absorber. The rich absorbent also passes through the heat exchanger before being pumped back to storage. The operation of the control system is intermittent, starting when the vapor holder is filled and stopping when it has emptied. Cleaned gases are vented from the absorber column to atmosphere.

A vapor recovery system employing a compression-refrigeration-condensation unit makes use of a vapor holder to store accumulated air-vapor mixture, and a gasoline saturator for ensuring that the VOC concentration is above the explosive range. The unit is activated and begins processing vapors when the vapor holder has filled to a preset level. Incoming saturated air-vapor mixture is first compressed in a two-stage compressor with an intercooler. Condensate is withdrawn from the intercooler prior to compression in the second stage. The compressed vapors then pass through the refrigeration-condensor section where they are returned along with the intercooler condensate to a gasoline storage tank. Cleaned gases are exhausted from the top of the condensor.

Costs for vapor processing units on fixed roof tanks vary with the type of product and the product throughput. In the case of gasoline, capital investment costs for these units range from $152,000 to $270,000 for a gasoline throughput of 380 m^3/day. These costs increase by 15% when the gasoline throughput increases by 150%. Costs for internal floaters in fixed roof tanks range from approximately $4,000 to $40,000 for storage tanks with diameters of 5 and 30 meters, respectively. For floating roof tanks, controls of evaporative emissions by the addition of secondary seals are estimated to cost $75 per linear meter.

Fugitive Organic Emissions from Process Equipment

There are many potential sources of fugitive organic emissions that result when process fluid (either liquid or gaseous) leaks from plant equipment in a typical gasification/liquefaction synthetic fuel plant. Some of these sources are: pumps, compressors, in-line process valves, pressure relief devices, open-ended valves, sampling connections, flanges, and drains.

There are two basic methods which have been used to control fugitive organic emissions from process equipment:

1) leak detection and repair methods, and
2) equipment specification.

Leak detection methods include individual component surveys, area (walk-through) surveys, and fixed point monitoring at specified intervals. Guidelines for the frequency of leak detection surveys in petroleum refineries have been suggested by EPA (52). However, different guidelines might have to be developed for direct liquefaction plants, due to differences in equipment and composition of leaks between these plants and refineries.

In an individual component survey each fugitive emission source (pump, valve, compressor, etc.) is checked for VOC leakage. The source may be checked for leakage by visual, audible, olfactory, soap bubble, or instrument techniques. Visual methods are particularly effective in locating liquid leaks. Escaping vapors from high pressure leaks can be audibly detected, and leaks of odorous materials may be detected by smelling the odor. Perhaps the best method of identifying leaks of VOC from equipment components is by using portable detection instruments. By sampling and analyzing the air in close proximity to the leak the hydrocarbon concentration of the sampled air can be determined. The leak rate from the source can be estimated since relationships exist between monitoring concentrations and mass emission rates.

Air Source Type 5
Fugitive Organic
Emissions

An area survey (also known as a walk-through survey) requires the use of a portable hydrocarbon detector and a strip chart recorder. The procedure involves carrying the instrument within one meter of the upwind and downwind sides of process equipment and associated fugitive emission sources. An increase in observed concentration indicates leaking fugitive emission sources. The instrument is then used for an individual component survey in the suspected leak area.

Fixed point monitors are automatic hydrocarbon sampling and analysis instruments positioned at various locations in the process unit. The instruments may sample the ambient air intermittently or continuously. Elevated hydrocarbon concentrations indicate a leaking component. As in the walk-through method, an individual component survey is required to identify the specific leaking component in the area. For this method, the portable hydrocarbon detector is also required.

Reduction of fugitive emissions from the identified leaking components is effected by repair methods. In many cases, perfect repair will not be achieved; however, effective repair can substantially reduce emissions from the leaking component. Some repairs might not be performed while the plant is on-stream. Also, both the speed with which repairs are made and the size of leak requiring repair are issues addressed in EPA's refinery VOC leak control techniques guidelines document (52). Typical repair methods employed on the various components are listed in Table 4-40.

The second method used to control fugitive emissions is by equipment specification. Typical equipment specifications used are listed in Table 4-41.

Costs for leak detection and repair methods will depend upon the frequency of leak detection surveys, the number of type of fixed point monitors used, and the complexity of the component undergoing repair. The major costs of leak detection, maintenance and repair methods are labor related. In the

TABLE 4-40. REPAIR METHODS FOR FUGITIVE EMISSIONS REDUCTION

Component	Repair Method
Pumps and compressors	Tighten packing gland Replace pump and compressor seals
Relief valves	Manual release of the valve may improve the seat seal Replace relief valve seats
In-line valves	Tighten packing gland Lubricate plug type valves Inject sealing fluid into process valves requiring repair Replace valve packing
Flanges	Replace flange gaskets

TABLE 4-41. EQUIPMENT DESIGN/MODIFICATIONS FOR FUGITIVE HYDROCARBON EMISSIONS CONTROL

Component	Specification
Pumps	- improve seal at the junction of moving shaft and stationary casing - use sealess pumps - use double mechanical seals - use closed vent systems around seal areas
Compressors	- improve seal at the junction of moving shaft and stationary casing - use double mechanical seals - use closed vent systems around seal areas
Pressure Relief Devices	- use rupture disks upstream from the safety/relief valve - use resilient seal or "o-ring" relief valves - use closed vent systems to transport valve discharge to control devices
Open-Ended Valves	- install a cap, plug, flange, or a second valve to the open end of the valve
In-Line Valves	- use diaphragm and bellows seal type valves
Drains	- covers - drain traps.

case of equipment specifications, costs will depend upon the component being specified. Typically, double mechanical seals in conventional refinery applications cost $815/pump (installed). Flush oil systems for double mechanical seals cost $1500/pump. These costs could be higher in direct liquefaction applications, due to the potential abrasiveness of some materials that will be pumped in direct liquefaction plants. Leak detection/repair and equipment specification techniques are discussed in further detail in Appendix A-18.

4.2.5.1 Evaporative Emissions from Product/By-Product Storage Emissions (Stream 751)

Various types of storage vessels are employed to store petrochemical products. The suitability of a specific tank design depends on the vapor pressure that the stored product exerts at ambient conditions and the storage desired.

According to Exxon's designs, naphtha from the EDS process will be stored in floating roof tanks, whereas fuel oil and phenol will be stored in fixed roof tanks. For naphtha, uncontrolled emissions were estimated by assuming that the external floating roof tanks are equipped with only primary seals. For fuel oil and phenol, uncontrolled emissions were estimated by assuming that there are no vapor recovery systems and no internal floating roofs installed on the fixed roof tanks. Uncontrolled emission estimates from these tanks have previously been discussed in Section 3.3.5.5.1.

Emissions from floating roof tanks consist primarily of standing storage losses and withdrawal losses, as discussed previously. These losses are greatly reduced by the addition of secondary seals, which is the most widely used approach for VOC control in existing floating roof tanks. The addition of secondary seals can reduce overall emissions from naphtha storage tanks by 95%, and is the control option discussed here. Estimated controlled organic emissions from naphtha storage are presented in Table 4-42.

Emissions from fixed roof tanks consist primarily of breathing losses and working losses, as discussed earlier in this section. These losses can be reduced by the installation of vapor recovery systems, thermal oxidation for emission control, or internal floating roofs. For purposes of this discussion, emissions from fixed roof tanks for fuel oil and phenol storage are controlled by installation of internal floating roofs with secondary seals. The overall control efficiencies of vapor recovery systems and thermal oxidation could be potentially higher than the installation of internal floating roofs, but insufficient information concerning the properties of blended fuel oil and

TABLE 4-42. STORAGE TANK EMISSION ESTIMATES

Product	No. of Tanks	Roof Type	Capacity per Tank (m^3)	Diameter (m)	Vapor Pressure kPa	Mass Emission Rate* Uncontrolled† kg/yr	Mass Emission Rate* Controlled‡ kg/yr	Avg. Control Efficiency
Base Case								
Naphtha	2	Floating	41,200	62.5	10	18,750	905	95
Blended Fuel Oil	2	Fixed	36,800	59.4	0.00028	153	31	80
Phenol	2	Fixed	730	12.2	0.14	196	78	60
MFS Case								
Naphtha	2	Floating	40,000	62.5	10	18,730	880	95
Blended Fuel Oil	2	Fixed	34,000	56.4	0.00028	140	28	80
Phenol	2	Fixed	730	12.2	0.14	196	78	60

*Calculations based on information contained in AP-42 (21).

†Uncontrolled floating roof tanks were assumed to have only primary seals. Uncontrolled fixed roof tanks have no vapor recovery system and no internal floating roof.

‡Controlled floating roof tanks were assumed to have both primary and secondary seals. Controlled fixed roof tanks were assumed to have internal floating roof covers with secondary seals.

phenol and site-specific factors was available to allow the design, evaluation and costing of these alternative control systems. As shown in Table 4-42, only approximately 15% of the estimated controlled evaporative organic emissions would be from fuel oil and phenol storage, after the installation of internal floating roofs with secondary seals.

In Table 4-43, the capital investment and annualized costs for control of evaporative emissions from storage of naphtha, fuel oil, and phenol are presented. For either the base case or the MFS case design, the total capital investment cost for these controls corresponds to only 0.008% of the uncontrolled base plant cost. The annualized control cost for either EDS design corresponds to less than 0.006% of the uncontrolled base plant cost.

TABLE 4-43. COSTS OF CONTROL OF EVAPORATIVE EMISSIONS FROM PRODUCT AND BY-PRODUCT STORAGE (STREAM 751)

Product	Capital Investment Cost*			Annualized Cost +		
	Total Cost, $ million	% of Base Plant	Unit Cost, $ per kg/yr product recovered	Total Cost, $ million	% of Base Plant	Unit Cost, $ per kg product recovered
Base Case						
Naphtha	0.051	0.002	2.9	0.012	0.001	0.67
Blended Fuel Oil	0.155	0.005	1,270	0.036	0.004	295
Phenol	0.030	0.001	387	0.007	0.0007	90
MFS Case						
Naphtha	0.051	0.002	2.9	0.012	0.001	0.67
Blended Fuel Oil	0.147	0.005	1,310	0.034	0.004	304
Phenol	0.030	0.001	389	0.007	0.0007	295

* The capital investment costs for control of evaporative emissions correspond to: 1) installation of secondary seals for naphtha storage, 2) installation of internal floating roofs with secondary seals for fuel oil and phenol storage. Costs were obtained from data presented in Reference 53.

+ The annualized costs were derived using methodology described in Section 4.1.3. The annualized cost for control of evaporative emissions from naphtha storage is reduced by $4,600 for products recovered. Credits for products recovered from control of evaporation emissions in fuel oil and phenol storage are negligible.

4.2.5.2 Fugitive Organic Emissions from Process Equipment

As discussed earlier, process equipment such as pumps, compressors, in line valves, pressure relief devices, open-ended valves, etc., are prone to leakage and thus are sources of fugitive organic emissions. Two methods can be employed to control these emissions. A labor intensive method involving leak detection and constant repair and maintenance can be used and/or equipment can be employed which is designed to reduce the risk of leaks. Obviously, if equipment specification in addition to extensive leak detection and repair is performed, greater reduction in fugitive organic emissions is achieved.

Two approaches to reduce fugitive organic emissions are generally used. In the first approach leak detection and repair methods as suggested in the VOC leak control techniques guideline document for the petroleum industry can be applied (52). Using this approach, controlled emission estimates for the EDS based direct liquefaction facility under consideration were made. An overall emission reduction of approximately 66% was estimated for leak detection/repair only, as shown in Table 4-44. These estimates were based on a detection level of 10,000 ppmv, weekly inspections of light liquid pump seals, monthly inspection of all other equipment, and open-ended valves were required to be sealed with a cap plug or another valve. Capital investment (TCI) and annualized costs for leak detection/repair only were estimated to be $39,000 and $65,000, respectively, using the labor requirements (Table A 18-7) and equipment cost information provided in Appendix A-18. These costs correspond to 0.001% of the base plant capital investment and 0.007% of the base plant annualized costs.

The second approach is a step beyond the first in that it relies on equipment specification in addition to leak detection, and repair and maintenance. Monitoring requirements are similar to those for the first approach, except in cases where equipment spcification eliminates the need for monitoring. On applying equipment spcifications in addition to leak detection/repairs to an

TABLE 4-44. FUGITIVE ORGANIC EMISSIONS FROM PROCESS EQUIPMENT FOR EDS COMMERCIAL PLANT (WITH LEAK DETECTION/REPAIR AND EQUIPMENT SPECIFICATION)

Components/ Service	Uncontrolled Emission Factor kg/hr/unit	Uncontrolled Emission Rates, kg/hr	Controlled Emission Rates,* kg/hr
Pump Seal			
Light Liquid Service	0.113	8.8	0-2.2
Heavy Liquid Service	0.021	3.4	0-2.9
In-Line Valves			
Gas Service	0.027	32.4	3.2
Light Liquid Service	0.011	19.5	5.1
Heavy Liquid Service	0.00023	0.8	0.8
Hydrogen	0.0082	4.2	0.4
Safety Relief Valves			
Vapor Service	0.086	21.7	0-8.3
Compressor Seal			
Hydrocarbon	0.64	36.2	0-10.9
Hydrogen	0.05	0.9	0-0.4
Flanges	0.00025	6.5	6.5
Drains			
Light Liquid	0.032	4.8	2.8
Heavy Liquid	0.045	13.6	7.8
Totals		153	26.6-51.3

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* The higher numbers in emission rates correspond to control by leak detection/repair methods only and the lower numbers correspond to control by both leak detection/repair and equipment specification. Where a single value is given, information on controlled emissions from equipment specification is not available.

EDS commercial plant, an overall emission reduction of approximately 83% can be achieved, with the combination of these two approaches, as shown in Table 4-44. The emission reduction was estimated on the basis of the following equipment specifications: 1) installation of caps for all open-ended valves, 2) use of double mechanical seals on pumps and compressors, 3) use of flush oil system for double mechanical seals, 4) installation of vents for compressor and pump degassing reservoirs, 5) installation of rupture disks for relief valves, and 6) use of closed loop sampling connectors. The capital investment costs for equipment specifications were calculated by multiplying the number of these components required and the cost per component provided in Appendix A-18 (2) (e.g., the capital investment cost for double mechanical seals on pumps was derived by multiplying the number of pump seals given in Table 3-73 by the installed cost (IEC) for each double mechanical seal), and added to the capital investment cost for leak detection/repair. The annualized cost for both leak detection/repair and equipment specification was calculated based on the labor requirements given in Appendix A-18 and the methodology described in Section 4.1.3. The capital investment (TCI) and annualized costs for these combined controls were estimated to be \$1.6 million and \$0.66 million, respectively. These costs correspond to 0.05% of the base plant capital investment and 0.07% of the annualized base plant cost.

Detailed bases for the emission reductions and costs are provided in Appendices A-18 (1) and A-18 (2).

4.2.6 Source Type 6 - Fugitive Particulates from Material Conveying and Processing

There is only one waste stream identified under Source Type 6:

- Stream 013 - fugitive particulates from coal handling and crushing.

Material transfer and conveying operations are common to nearly all processing industries. Equipment includes belt conveyors, screw conveyors, bucket elevators, vibrating conveyors, and pneumatic conveyors. The type of conveying equipment varies with the application, and is determined primarily by the quantity and characteristics (size, specific gravity, moisture content, etc.) of the material being handled, the transfer distance and elevation, and conditions of the working environment. Loss of material from conveyors is primarily at the feeding, transfer, and discharge points and occurs due to spillage or windage. The majority of particulate emissions are generally from spillage and mechanical agitation of the material at transfer points.

Coal from storage piles is generally crushed, screened, and pulverized prior to transfer to the boiler or EDS liquefaction area. Fugitive dust generated during this crushing and handling process is typically controlled by either wet suppression techniques or dry particulate collection systems.

Wet suppression systems utilizing a wetting agent consist of pre-engineered modules which incorporate both water handling components and automatic spray controls. A typical spray solution contains 1,000 to 4,000 parts of water to one part of a wetting agent. The rate of spray application is about 4 to 8 liters/Mg of material. This rate of application results in an increase of total surface moisture by about 0.5 to 1.0%.

In wet dust suppression the fugitive particulate is first confined by a curtain of moisture droplets. Then the wetting of dust takes place by contact and penetration with moisture droplets. Finally, agglomerates are formed by contact with other droplets and settling takes place because of the additional weight of the other droplets. Wet suppression techniques can cost from $0.33 to $0.77/Mg, depending upon the wetting agent utilized.

Dry particulate collection systems consist of enclosures to contain the particulates, ductwork and exhaust systems to convey the contained particulate laden air, and particulate collectors to separate the particulate from the air. Typically, exhaust hoods can be used to capture particulate emissions at transfer points. Conveyors generally have a half cover which provides dust containment and also shields the conveyor from wind, rain and snow. Enclosure sizing is a function of the source under consideration. The type of enclosure used depends upon the particulate source. Typical ductwork velocities in exhaust hoods for particulate capture for different source types are readily available. Dust collectors that are applicable to the collection of the captures particulate are: 1) venturi scrubbers, 2) electrostatic precipitators, 3) fabric filters, and 4) dry centrifugal collectors. These have been discussed previously in Section 4.2.2.

4.3 WATER POLLUTION CONTROL TECHNOLOGIES

The EDS process generates many wastewater streams which may need various forms and degrees of control prior to discharge or reuse. The commercially available techniques that are potentially applicable to the EDS wastewater streams are identified and evaluated in this section. The primary focus of the material presented is on applying the control techniques to streams specific to the EDS process.

A conclusive and concise selection of one control technique over another would require more data than is currently available as well as an in-depth analysis of many site- and design-specific tradeoffs. Performing such an analysis is outside of the scope of this manual. In reviewing applications for specific proposed new facilities, permit reviewers are encouraged to look carefully for evidence that proposed controls are supported by adequate treatability and design data.

The applicability, performance, cost, reliability and other process related data reported here are based on engineering judgement, available literature data, and specifications and costs supplied by process vendors or licensors. Areas of particular concern with respect to many of the control techniques likely to be used in an EDS facility are identified throughout this section.

The presentation in this section is organized as follows. The waste streams characterized in Section 3 are divided into two source types, based on their characteristics and treatment requirements. The two source types are:

- Source Type 1 - Organic and Dissolved Gas-Containing Wastewaters
- Source Type 2 - Inorganic Containing Wastewaters

Under each source type, the generic approach in treating the streams is presented, followed by an evaluation of the applicable control techniques when applied to each individual stream. Some integrated control examples are then discussed to illustrate possible ways of combining the individual control techniques.

4.3.1 Source Type 1: Organic- and Dissolved Gas-Containing Wastewaters

Streams that fall into this source type include: a) streams that are common to both the Base and MFS Cases.

Stream No.	Stream Description
103	Slurry dryer cold separator wastewater
106	Liquefaction cold separator wastewater
152	Atmospheric fractionator overhead drum wastewater
155	Vacuum fractionator overhead drum wastewater
202	Solvent hydrogenation cold separator wastewater
252	Solvent hydrogenation fractionator overhead drum wastewater
308	Flexicoking fractionator overhead drum wastewater
307	Flexicoking recontacting drum wastewater
312	Flexicoking heater overhead drum wastewater
403	Knockout drum wastewater in H_2 cryo recovery

b) streams that are unique to the Base Case:

Stream No.	Stream Description
430	Blowdown and K.O. drum wastewater from H_2 generation
431	Overhead receiver wastewater for AGR in H_2 generation
452	Knockout drum wastewater in ammonia synthesis
451	Aqueous ammonia from ammonia synthesis

c) streams that are unique to the MFS case:

Stream No.	Stream Description
441	Sour water from partial oxidation unit
443	Slag filtrate from partial oxidation unit

In addition to these streams, certain secondary waste streams which also belong to this source type will exist in a controlled EDS plant. Examples of these include Beavon Sour Condensate and SCOT Sour Condensate. These secondary streams are identified where the appropriate control techniques are discussed (e.g., Section 4.2.1.1.2.1 for Beavon Sour Condensate) and they could be treated along with the streams listed here.

As discussed in Section 3, these streams contain, among other pollutants, high levels of phenols and/or dissolved gases such as ammonia and hydrogen sulfide. Controlling these streams typically involves more than one technology arranged in series. Table 4-45 presents some summary information about the commercially available treatment techniques that are potentially applicable to these waste streams. While this list of control techniques is believed to be fairly complete, it is by no means inclusive; other techniques may be available or may be developed which could also be applicable. The techniques presented in Table 4-45 are broadly divided into seven categories according to their functions within the overall control sequence. The seven control areas are:

1) suspended solids, tars and oils removal
2) bulk organics removal
3) dissolved gases removal
4) dissolved organics removal
5) residual organics removal
6) volume reduction, and
7) final disposal.

Table 4-45. SUMMARY OF CONTROL TECHNIQUES POTENTIALLY APPLICABLE TO THE TREATMENT OF EDS WASTEWATER CONTAINING PRIMARY DISSOLVED ORGANICS AND DISSOLVED GASES

Technology	Technology Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-Products and Secondary Waste Streams	Comments
Removal of Suspended Solids, Tars and Oils						
Gravity separation - API oil separator - Parallel plate	Provision of adequate residence time in a stagnant vessel to allow suspended solids or immiscible fluids to separate into lighter and heavier than water phases.	Suspended solids, tars and oils.	Depends upon design, 10-50% removal of TSS typical, 60-99% for oils.	Minimum feed stream turbulence.	Recovered oils (Sp. gr. < 1), sludges/solids (Sp. gr. >1)	Incorporated into the tar/oil separation system design and the biological treatment system.
Coagulation/ flocculation	Use of agents to promote the coalescence of fine suspended solids and adsorption of tars and oils; generally used in conjunction with a gravity separation process.	Promotes removal of finely dispersed particles.	Outlet suspended solids concentration to 10 mg/l possible, oils removal of 60-95%.	A wide range of commercial flocculants are available.	Same as gravity separation.	Widely used in water treatment system to remove fine solids.
Air flotation - Dissolved air - Induced air	Use of air bubbles to promote the disengagement of lighter-than-water materials from solution.	Suspended oils and solids.	Depends on characteristics of source and treatment process; TSS removal of 20-75%, oil removals of 75-85%.	Air requirements depend upon waste characteristics.	Recovered oils entrained, solids.	Widely used in various industries.
Filtration	Passing wastewater through suitable filter medium; filter material discarded or cleaned by backflushing.	Depends on filter medium; both coarse and fine structure materials are used industrially.	TSS removals of 30-90+%, oil removals of 65-90%	Filter media (sand, clay, fabric or polymeric membrane).	Filter backwash; spent filter media.	Proposed for use as polishing step for phenolic water downstream of tar/oil separation; "Sticky" tars/oils may cause problems with filter plugging and regeneration.
Removal of Bulk Organics						
Solvent extraction - Phenosolvan - Chem-Pro	Extraction of organics from wastewater via contact with an immiscible solvent. By-product organic liquids recovered from the solvent in a separate regeneration step.	Phenols, TOC, BOD, COD other organics.	Phenosolvan: - monohydric phenol 99.5% - polyhydric phenol 60% - organic acids 15% Chem-Pro: - monohydric phenol 99.8% - dihydric phenol 95% - trihydric phenol 90-95% - other organics 50%	Sensitive to suspended matter in wastewater, makeup solvent.	Crude phenols, filter backwash, spent filter media.	Commercially proved in coke-oven industry and Lurgi gasification plants. First plant installed in 1940. Several commercial applications in phenol formaldehyde resin manufacturing plants. First application in 1961 to recover phenol from coke-oven waste liquor.
Resin Adsorption	Selective removal of phenols via contact with a solid medium (the resins).	Phenols, TOC, BOD, COD other organics.	90+% removal of phenols.	Acetone for regenerating the resin.	Crude phenols.	Commercially proven in organic chemical industry.

(Continued)

TABLE 4-45. (CONTINUED)

Technology	Technology Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-Products and Secondary Waste Streams	Estimated Capital Cost	Comments
Removal of Dissolved Organics							
Biological oxidation - act. sludge - trickling filter - rotating biological contactor - lagoons - high purity oxygen (HPSAS)	Biological conversion of the carbonaceous organic matter in wastewater to cell tissue and various gaseous end products.	TOC, BOD, COD, some inorganic pollutants	Varies with waste stream characteristics. Typical removals: BOD 60-90 COD, TOC 80 total phenol 95 org acids 95 oil & grease 70 CN⁻ 70	Air or oxygen, supplemental nutrients may be required; relatively constant feed temperature and pollutant loadings required to minimize "shocks" to system	Biological oxidation sludge	Activated sludge: $700 per kg BOD per day	This is the basis for the treatment of coke oven wastewaters
Wet air oxidation	Direct reaction of O_2 with wastewater in a closed, pressurized vessel at elevated temperatures	Same as biological oxidation except better destruction of cyanides and other difficult to treat organics can be achieved	Over 90% removal of COD is possible in a system with a residence time of 1 hour or greater. Actual efficiency heavily dependent upon waste characteristics.	Air or oxygen; heat if autothermic reaction conditions are not present	Vent gases containing CO, CO_2, light hydrocarbons, NH_3, sulfur species	--	Promising but not proven in this application; fairly rigorous materials of construction requirements.
Ultrafiltration	Forcing wastewater through semipermeable membrane under pressure	Most effective with high molecular weight organics	Up to 95 removal for total organics	Filter media	Spent filter media; concentrated wastewater	--	May be attractive as a preconcentration step prior to wastewater incineration.
Anaerobic digestion	Reduction of organics in closed vessel at moderate temperatures to form CH_4 and other hydrocarbons, digestion process relies upon metabolic processes of anaerobic organisms	TOC, BOD, COD	Unknown - this process has not been applied to these types of waste streams	Some supplemental nutrients may be required	Waste gases (combustible)	--	Kinetic limitations and process control problems could be substantial.

(Continued)

TABLE 4-45. (CONTINUED)

Technology	Technology Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-Products and Secondary Waste Streams	Comments
Removal of Dissolved Gases						
Steam stripping	Increasing temperature and providing a positive flow of inert material (steam) through the wastewater; removes volatile organics and inorganics with overhead steam.	NH_3, acid gases (CO_2, H_2S, HCN), light hydrocarbons (phenols).	95-99% removal of "free" ammonia and acid gases typical, hydrocarbon removal varies with volatility of stripped components.	Feed preheat can be used to reduce steam requirements; acid/caustic for pH adjustment optional.	Stripped gases, uncondensed steam.	Acid/caustic addition can be used to improve the efficiency of the stripping process.
- Phosam-W	Volatile contaminants are stripped and NH_3 absorbed by phosphate salts, NH_3 recovered; scrubbed sour gases to sulfur recovery plant.		Removal to: 150 mg/ℓ NH_3 1 mg/ℓ H_2S, CO_2, 95% HCN.			
- Chevron WWT	Volatile contaminants are stripped in two stages with the H_2S and CO_2 removed in the first stage, NH_3 removed and subsequently recovered in the second stage.		Removal to: 50 mg/ℓ NH_3, 5 mg/ℓ H_2S.			
Inert gas stripping	Same as with steam except that an inert gas such as N_2, air or CO_2 is used as the stripping medium.	Same as for steam stripping.	Same as for steam stripping.	Generally operates at lower temperature than steam stripping process.	Stripped gases, stripping gas.	The presence of an inert stripping gas as a diluent may make it more difficult to handle or further treat the gases stripped from the wastewater.
Vacuum distillation	Low pressure, low temperature stripping process.	Same as for steam stripping.	Same as for steam stripping.	Same as for inert gas stripping.	Stripped gases.	High energy requirements; not cost competitive in a plant where stripping steam is readily available.
Selective absorption	Used in conjunction with one of the above processes to selectively absorb stripped gases.	NH_3 (with acidic solutions), acid gases (with basic solutions).	90% removal of acid gases or NH_3 is typical.	Makeup acid/caustic.	Recovered NH_3 or acid gases, rich sorbent.	This is the basis for several commercially proven processes for recovering high quality ammonia.
Removal of Residual Organics						
Activated carbon adsorption	Adsorption of organics in water by activated carbon or polymeric resin. Powdered activated carbon has been used in conjunction with biological processes (above) with some success in the organic chemical industry.	Most effective with complex organics and phenols; some heavy metal removal expected.	Varies with wastestream characteristics. Typical removals: BOD 60 COD 80 TOC 70 Phenols 99.9 org acids 70 CN 50 SCN^- 50 Tars 99 oils 99	Adsorbent.	Spent adsorbent regeneration off-gases.	Probably more effective as a polishing rather than a bulk organics removal process.
Chemical oxidation	Reaction of organics in wastewater with ozone, peroxides or chlorine-based oxidants.	TOC, BOD, COD, oxidizable inorganics.	High removals achievable depending upon conditions of operation.	Oxidant.	Vent gases, wastewater and reaction products.	Chlorine-based oxidants may cause problems with treated wastewater.

(Continued)

TABLE 4-45. (CONTINUED)

Technology	Technology Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-Products and Secondary Waste Streams	Comments
Thermal oxidation (Incineration)	Combustion of organics.	All oxidizable organics including complex organics such as PAH's.	Essentially complete destruction of organics in properly designed system.	Supplemental fuel, preconcentration will improve performance and lower supplemental fuel requirements.	Flue gases.	This will be the most effective process for removing organics but the supplemental fuel requirements may be substantial.
Cooling tower oxidation	Air oxidation (and stripping) or organics and dissolved gases.	TOC, COD, BOD, phenols and organics, NH_3.	Unknown - this process has been tested using SASOL wastewaters on a pilot scale but results are not available.	Sensitive to suspended matter, oils, high TDS.	Blowdown/drift.	Treated wastewater use in small refinery cooling towers has been practiced - 75-80 reduction in wastewater volume is common.
Volume Reduction						
Membrane separation - Reverse Osmosis	Use of semipermeable membrane and pressure to separate water from its dissolved constituents.	Relative rejection efficiencies of the various soluble species will be determined by membrane characteristics and conditions of process operation.	90-95 rejection of dissolved salts. Reductions in dissolved organics and BOD of up to 99 .	Membrane.	Spent membrane material, recovered water, brine.	May be useful as preconcentration step prior to further treatment or ultimate disposal of wastewater. Membrane scaling and fouling with organics may limit the applicability of this technology to the treatment of process condensates.
- Electrodialysis	Use of selective anion- or cation-permeable membranes with electric field to separate mineral ions from water.					
Forced evaporation	Thermally induced evaporation of excess wastewater, condensate recovery optional.	All nonvolatile species will remain in brine.		None	Recovered condensate, noncondensible gases, waste brine.	Very stringent materials requirements; **significant energy requirements**.
Cooling tower concentration	wastewater used as partial makeup to the cooling tower and thereby concentrated into the blowdown.	All nonvolatile species concentrated into the blowdown.		Feed characteristics limited by corrosion, scaling, and biological fouling.	Evaporation/drift.	
Residual Disposal						
Surface discharge	Wastes are conveyed to and mixed with a natural water body.	Entire stream.	100	Restrictions are site-specific **depending on local stream background levels and applicable water quality requirements**.	None	

(Continued)

TABLE 4-45. (CONTINUED)

Technology	Technology Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-Products and Secondary Waste Streams	Comments
Deep well injection	Wastes are pumped into subsurface geological formations where they are isolated from all surface and groundwater supplies.	Entire stream.	100	Injected fluids must be filtered to 5 micrometers and have a low organic content to prevent plugging. Wastes must not precipitate in the well or when mixed with subsurface fluids. Volume reduction prior to injection is often economical.	None	**Applicability of this technique depends on the availability of suitable subsurface formation.**
Surface impoundment	Wastes are held in a containment basin.	All nonvolatile species remain as residual in the impoundment.	100	Concentrations of volatile species may need to be low to prevent loss to the atmosphere.	Loss of volatile species.	This technique is limited to locations having available land and net evaporation rates exceeding 0.5 m/yr (20 in/yr).
Co-disposal	Ash is quenched with wastewater then handled by a solid waste disposal technique.	Entire stream.	100	Concentrations of volatile species may need to be low to prevent loss to the atmosphere.	Loss of volatile species.	

Although an integrated system for controlling some streams may often require one technique from each of the functional areas arranged in the same sequence as listed, this is not always the case. Depending on the characteristics of the stream and the control requirements, an integrated control scheme may include individual techniques from only a few of the areas and/or may suggest more than one individual technique from the same functional area. Coagulation/flocculation + Gravity Separation and Chemical Oxidation (ozonation) + Carbon Adsorption are good examples where more than one technique from the same area might be combined in a treatment scheme.

The removal efficiency data presented in Table 4-45 are "typical optimum" performances reported for industrial/municipal applications and are believed to be reasonable estimates for the EDS wastewaters. In principle, any control technique can be designed to remove the pollutants it is capable of removing to any degree of efficiency. Likewise, a treatment scheme can be selected to control a stream to any degree. However, primarily due to economic constraints, a typical, or optimum removal efficiency is often used by process developers/designers, and this may vary depending on the waste stream characteristics and site-specific factors. For the same reasons techniques in different functional areas are usually combined in typical arrangement; all techniques in one functional area are not necessarily candidates for integration with each technique in another area.

Control Function 1 includes those techniques that are capable of removing tars and oils, and suspended solids. These materials may plug up other control equipment or interfere with the control mechanisms. Thus, to protect the downstream control techniques or improve their efficiencies, Function 1 is generally placed ahead of other processes. For costing purposes, this function is considered as part of the EDS process. The oils and tars removed are recycled back to the process.

Water Source Type 1

Control Function 2 includes those techniques that are applicable to treating wastewater with high phenol levels. This function may be placed ahead or after Function 3 - Dissolved Gases Removal. Reversing the sequencing of these two functions may generate different secondary waste streams. If Function 2 technique is placed ahead of Function 3, some of the dissolved gases will be stripped during the residual solvent recovery step of the Function 2 extraction processes. If the sequence is reversed, some phenols and other volatile organics will be stripped by the stripping step in Function 3. In either approach, these components are recovered by minor adjustment of the auxiliary equipment. The stripped dissolved gases may be routed to the ammonia recovery unit to recover the ammonia; the stripped phenol may be recovered in a condenser.

Due to the high levels of phenolics and dissolved gases in some EDS wastewaters, they are good candidates for removal before other conventional treatment techniques are applied. Phenol removal techniques include solvent extraction, resin adsorption and oxidation techniques. Both solvent extraction and resin adsorption generally recover the phenolics in the waste stream while oxidation destroys the phenolics. The solvent extraction and adsorption techniques are expected to recover more than 95% of the phenolics in the EDS wastewaters, and oxidation can destroy more than 90% of the organics. Dissolved gases removal techniques generally employ a steam stripping step to remove dissolved gases such as CO_2, H_2S and NH_3, and an absorption step to recover ammonia from the stripped gases as by-product. These techniques are expected to remove more than 99% of the CO_2 and H_2S, recover the bulk of the ammonia, and generate a stripper bottom stream with about 50 to 200 ppm NH_3.

After the phenolics and dissolved gases are removed, other dissolved organics can be destroyed by one of the dissolved organics removal techniques (control function 4). These techniques include biological oxidation (biox) which involves the use of microorganisms to breakdown the organic molecules; other dissolved organics removal techniques include wet air oxidation.

Typically, biox techniques remove about 80% COD and TOC.

Biological processes are very sensitive to fluctuation in the incoming wastewater characteristics such as pH, temperature, toxic chemicals, nutrients, etc. Significant changes in any of these parameters could reduce the overall treatment efficiency or even result in a complete shutdown of the system due to destruction of microorganisms. Provision of adequate holding capacity ahead of the bio-treatment system would reduce the effect of fluctuation in waste characteristics and is a must in integrated treatment facilities. The toxic and the synergistic effects of various pollutants on the biogrowth are not completely known. However, raw EDS waste streams contain such high concentrations of phenol, ammonia, and hydrogen sulfide that they are likely to be toxic. Thus, pretreatments to remove these constituents are required to ensure proper operation of the bio system. In other words, Functions 2 and 3 would generally be placed ahead of Function 4 for those EDS streams high in phenols and dissolved gases.

The residual organics removal techniques (Control Function 5) are either physical sorption or chemical oxidation processes. These are considered polishing steps and can remove 90% or more of the residual organics, depending on the organic characteristics and process designs.

Volume reduction (Control Function 6) involves reducing the volume of wastewaters that require further treatment or disposal. This is usually achieved by either evaporating the water or concentrating the pollutants into a much smaller volume by physicochemical means. A greater than 90% reduction in waste volume can be achieved, depending on process selection and design. This function is generally considered only for integrated wastewater systems involving zero discharge or recovery and reuse of the wastewater.

Final disposal (Control Function 7) includes fundamentally different disposal approaches which may be feasible for use at a given site, depending on site-specific factors such as the availability of land, net evaporation rates, groundwater reservoir characteristics, and surface discharge limitations. The choice of disposal techniques will have profound impacts on the treatment techniques selected for treating the waste streams since each disposal alternative requires different levels of treatment to render the wastewater amenable to disposal, reuse or discharge. For example, surface discharge of treated effluent may involve a combination of techniques from Function 1 through Function 5. If zero discharge is employed, the volume of the treated effluent from Function 5 technique may be reduced using Function 6 technique. The brine thus generated may be disposed in deepwells, surface impoundments (solar evaporation pond) or co-disposed with solid wastes.

The application of specific control techniques to the individual EDS waste streams under Source Type 1 is discussed in the next section, followed by some integrated control examples.

4.3.1.1 Combined Stream A - Combined Sour Water Stream

Combined Stream A consists of seven sour water streams which are present both in the base case and the MFS case. These streams are:

- Stream 106 - Liquid Cold Separator Wastewater
- Stream 152 - Atmospheric Fractionator Overhead Drum Wastewater
- Stream 155 - Vacuum Fractionator Overhead Drum Wastewater
- Stream 202 - Solvent Hydrogenation Cold Separator Wastewater
- Stream 252 - Solvent Hydrogenation Fractionator Overhead Drum Wastewater
- Stream 307 - Flexicoking Recontacting Drum Wastewater
- Stream 308 - Flexicoking Fractionator Overhead Drum Wastewater.

These streams are generally combined for treatment because they all contain high concentrations of H_2S, NH_3, CO_2 and phenols.

Table 4-46 presents the estimated characteristics of the combined stream; the characteristics of the individual streams have been discussed in Section 3. Major species for which data/estimates are available for the seven individual streams listed above are included in the Table. As presented in Section 3, more than 10 classes of organic compounds have been quantified in four of the streams from pilot plant sampling (37). However, there are insufficient data to allow a reliable estimation regarding the levels of these trace species that would consistantly be observed in commercial EDS streams.

With the exception of COD, TOC, TDS, and Oil and Grease, all data presented in Table 4-46 are based upon Exxon design estimates (13). No COD, TOC, TDS or Oil and Grease data are estimated in the Exxon design, but they were determined for Streams 106, 152, 155 and 252 in pilot plant sampling by EPA (37) and Exxon (23). TOC and COD values in Combined Stream A are estimated by using the data on the 4 of the 7 component streams for which pilot data are available and assuming TOC and COD in the other 3 missing streams (202, 307 and 308) are present in the same ratio to phenol in these 3 streams as in the other 4 streams

TABLE 4-46. ESTIMATED CHARACTERISTICS OF STREAM A - COMBINED SOUR WATER STREAM (1,134 Mg/hr Coal Feed EDS Plant)

Components	Concentration*, mg/l Base Case	 MFS Case
H_2S	18,000	22,000
NH_3	14,000	17,000
CO_2	8,700	10,000
Phenols	9,300	11,000
Organic Acid	3,200	4,000
COD	45,000	51,000
TOC	14,000	16,000
Oil and Grease	200	250
TDS	4,000	4,900
Flow Rate, m^3/hr	260	210

* All values except those for TDS, COD, TOC and Oil and Grease are based on Exxon Design Estimates (13). TDS, COD, TOC and Oil and Grease values are prorated from EPA and Exxon test results (23,37), and Exxon Design estimates.

sampled at the pilot plant. The TDS and Oil and Grease values in Combined Stream A are estimated assuming their values in the 3 missing streams are negligible.

4.3.1.1.1 Control Function 1 - Suspended Solids, Tars and Oil Removal

The combined waste stream will have about 200 ppm to 250 ppm of tars and oils which may need to be removed prior to other treatment to ensure efficient operation. Several techniques discussed below can be applied to achieve this.

4.3.1.1.1.1 Control Technique 1 - Gravity Separation

This technique operates under the principle that by providing enough residence time in a quiescent reactor, the tars and oils will be separated from the wastewater due to their difference in density. For rectangular or circular type of separators wastewater is introduced into one end or the center of the separator, flowed through a diffusion baffle and the length of the separator. Oils and tars, which float to the surface, are pushed into slotted pipes by flight scrapers. At the same time the flight scrapers push any sludge deposits on the bottom of the reactor to sludge hoppers. For tilted parallel plate type separators, the oil floats to the top of the plates (or tubes) while the treated wastewater exits through the bottom of the plates. See Appendix B-1 for more detailed discussion of gravity separation. Based upon related experience in refinery operations (54,55), this technique is expected to remove 40 to 80% of the oils and tars in the Combined Stream A. For treating a 260 m^3/hr combined stream, the total annualized cost is estimated to be 1.3¢/m^3 (5¢/1000 gal.) (56).

Secondary Streams

In refinery and other industrial operations, gravity separation will generate two secondary waste streams, namely, an oils and tars stream and a settled sludge stream. However for Combined Stream A treatment, it is likely that minimal or no settled sludge will be formed because the stream contains only small amounts of suspended solids (less than 20 ppm). Thus only one secondary waste stream is expected from this operation. However, the tars present may have a density greater than water. Thus these systems should be designed to handle settled tars. Assuming a 75% removal efficiency, and that the oil skim pipes removes four times as much water as oil, the secondary tars and oils stream generated will amount to 330 l/min (88 gal/min). This stream can be incinerated or disposed with other wastewater treatment sludges. See Appendix B-1 for a more detailed discussion of gravity separation.

4.3.1.1.1.2 Control Technique 2 - Air Flotation

In air flotation, the removal of oil is achieved by dissolving air in the wastewater under pressure and then releasing the air at atmospheric pressure in a flotation tank. The released air forms tiny bubbles which adhere to the oil globules surface, causing the oil to float to the surface where it is removed by skimming. Flocculant aides such as polyelectrolytes, alum, ferrous sulfate and lime are commonly added to improve the flotation process. Oil removal efficiency of greater than 80% can be achieved with this technique (57). For treating Combined Stream A, the total annualized cost is estimated to be 0.3¢/m^3 (1.1¢/1000 gal.). See Appendix B-3 for bases and assumptions for the cost estimates.

Secondary Stream

One secondary waste stream is generated from the air flotation operation, namely, the recovered oil and tar stream. Assuming a 75% removal efficiency and that four times as much water as oil, this secondary stream generated will amount to 330 l/min (88 gal/min). This stream can be incinerated or disposed with other wastewater treatment sludges.

4.3.1.1.1.3 Control Technique 3 - Filtration

In filtration, wastewater enters the filtration unit and slowly percolates through the filter media to the underdrain. Oil and other suspended matters are trapped in the filter media and are later flushed out of the filter during the backwash cycle.

In refinery operations, where the wastewater contains high levels of free oils, filtration is generally applied as a secondary treatment, after gravity-settling. However this may be used as primary treatment for the subject stream since the stream contains relatively low suspended solids and oils. Based on experiences from refinery operations, about 66 to 90% of the oil is expected to be removed by filtration (56,57). The capital investment for treating a 260 m^3/hr Combined Stream A is estimated to be $470,000,or less than 0.01% of the uncontrolled base plant cost (see Figure B-4-1 in Appendix B-4 for installed equipment cost and Table 4-1 for bases for converting installed equipment cost to capital investment). The total annualized cost is estimated to be $130,000/yr, or about 1% of the uncontrolled base plant cost.

Secondary Stream

Filtration generates one secondary waste stream, namely, the filter backwash. There are no data to estimate the characteristics of this stream. This stream is generally recycled to sedimentation unit for treatment.

4.3.1.1.2 Control Function 2 - Bulk Organics Removal

The Combined Sour Water Stream contains high levels of soluble organics, the bulk of which are phenolics (around 8,000 ppm to 10,000 ppm). The effect of the Control Function 1 (Suspended Solids, Tars and Oil Removal) techniques on these soluble organics is assumed to be minimal. The control techniques most appropriate for treating this stream are solvent extraction and resin adsorption processes which recover the phenolics, and oxidation processes which destroy the phenolics and other COD exerting materials. It has been reported that it is cost effective to recover phenol from wastewaters containing more than 2,000 ppm phenol (58). A study comparing applying solvent extraction, adsorption and biological oxidation to phenolic waste similar to the Combined Sour Waste Stream indicated that solvent extraction is the most economical of all three techniques (59). The following sub-sections evaluate the applicabilities of two extraction processes (Phenosolvan and Chem-Pro), one resin adsorption process (Rohm and Haas), and one oxidation process (Wet Air Oxidation). As discussed before, these techniques can be placed ahead of or after the Control Function 3 techniques.

4.3.1.1.2.1 Control Technique 1 - Phenosolvan Process

The Phenosolvan Process is a proprietary solvent extraction process developed by the Lurgi Company. In this process the waste stream is filtered with a sand filter and is then fed to a series of mixer-settlers where it contacts a lean organic solvent (such as butyl acetate, isopropyl ether, and Phenisol, a Lurgi proprietary solvent) (60) in countercurrent flow. The solvent extracts the phenols from the wastewater, and after solvent-water phase separation, the rich solvent is sent to a distillation column for solvent and phenol separation. The extracted wastewater is stripped with nitrogen in a packed tower to recover residual solvent and is then sent to subsequent treatment.

The performance of the Phenosolvan process is dependent upon a number of factors such as the characteristics of the wastewater, the number of extraction stages and the solvent-to-wastewater ratio used. It has been estimated that the process can remove 99.5% monohydric phenol, 60% polyhydric phenols and 15% of other organics (61). The breakdown of the phenolics in an EDS Combined waste stream sample is about 90% monohydric and 10% dihydric (62). Thus, it is expected that about 95% of the total phenols in EDS can be removed (see Table 4-47). This is consistent with performance of a Phenosolvan process operating on Lurgi gasification wastewater (63). A more recent Phenosolvan design recovers the monohydric and polyhydric phenols in separate stages (64). This design, when applied to the EDS wastewater, is expected to improve the overall

TABLE 4-47. ESTIMATED COMPOSITION OF FEED AND TREATED EFFLUENT FROM PHENOSOLVAN PROCESS (STREAM A)

	Base Case		MFS Case	
Component	Combined Stream Feed	Phenosolvan Effluent	Combined Stream Feed	Phenosolvan Effluent
Phenols, mg/l	9,300	460	11,000	550
TOC, mg/l	14,000	4,200	16,000	4,800
COD, mg/l	45,000	18,000	51,000	20,000
Organic Acids, mg/l	3,200	2,700	4,000	3,400
Flow, m^3/hr	260	260	210	210

Note: The single numbers (versus ranges) presented in this table represent a specific set of design and operating conditions for the EDS plant and for the control technique. Both cost and performance figures could vary due to uncertainties in cost and performance estimates and due to variations in design/operation of the EDS plant and of the control technique. This note is equally applicable to other stream-specific performance and cost tables in this section.

TABLE 4-48. SUMMARY OF COST ESTIMATES FOR PHENOSOLVAN PROCESS* (STREAM A) (1,134 Mg/hr Coal Feed EDS Plant)

Cost Item	Base Case	MFS Case
Capital Investment	$\$8.96 \times 10^6$	$\$7.62 \times 10^6$
Total Annualized Cost	$\$2.1 \times 10^6$	$\$1.7 \times 10^6$
% Uncontrolled Base Plant Capital Cost‡	0.28	0.26
% Uncontrolled Base Plant Annual Cost‡	0.22	0.19
Unit Capital Investment	\$35,000/m^3/hr	\$36,000/m^3/hr
Unit Annualized Cost	\$1.1/m^3	\$1.0/m^3

* See Appendix B-5 for detailed basis and assumptions

‡ Cost estimates for uncontrolled base plant are presented in Section 2.3.

phenol removal efficiency to at least 98% (64). The only available data on the overall bulk organic removal efficiencies of the Phenosolvan process are obtained from Lurgi gasification operations. It is known that the amount of organics removed is dependent upon the types of organics present. As presented in Section 3, portions of the organics in Combined Stream A have been quantified. However, data on the Phenosolvan removal efficiencies for these specific compounds are not available. Also, not all of the organics in the stream have been identified. For lack of a better data base it is assumed that the organic removal efficiencies for Combined Stream A are similar to those of the Lurgi waste streams, i.e., 60% for COD and 70% for TOC (65).

The estimated capital and annualized costs for this process are summarized in Table 4-48. The capital investment (TCI) shown in Table 4-48 is obtained by multiplying the installed cost (IEC) by 1.8139. This factor includes contractor overhead and profit, engineering and construction cost, contingency and interest during construction (see Section 4.1 for details). The installed cost is factored from data reported in the literature (see Figure B-5-2 of Appendix B-5).

The capital investments for the base and MFS cases are 8.96 and 7.62 million dollars, or about 0.28% to 0.26% of the base plant costs, respectively. The total annualized costs for the base case and MFS case are estimated to be 3.4 and 2.9 million dollars, respectively, or about \$1.1 m^3 and \$1.0/m^3 of Stream A flow: these figures were derived using the procedures summarized in Table 4-2, and the utility/solvent requirements presented in Appendix B-5. In estimating the annualized cost, recovered phenol is credited with its heating value only (2.2¢/kg), which is much lower than its potential resale value (20¢/kg). Thus, if there is a market for the recovered phenol, the annualized cost for this process could be considerably lower. For 20¢/kg, the base and the MFS cases will generate a profit of 1.1 and 0.9 million dollars per year, respectively.

Secondary Streams

The Phenosolvan process generates two waste streams: the treated effluent and the recovered phenol. The estimated characteristics of the treated effluent are presented in Table 4-47. It is not possible to estimate the characteristics of the recovered phenol because of the potential presence of other organics. Nonetheless, based on the removal efficiency, it is expected that the relative breakdown among monohydric phenol/polyhydric phenol/organic acid in the combined stream for the base case and MFS case will be 89%/6%/5% and 88%/6%/6%, respectively.

4.3.1.1.2.2 Control Technique 2: Chem-Pro Process

The Chem-Pro dephenolization process is a proprietary process developed by Jones and Laughlin Corporation to recover phenols from coke-oven waste liquors. Although the process is similar in many respects to the Phenosolvan process, this process uses a vertical reciprocating plate extractor column rather than a horizontal mixer-settler vessel to extract the phenols from the wastewater.

In this process, the wastewater is pumped to the top of the extraction column and the proprietary solvent is introduced to the bottom of the column. As the wastewater descends the column, it mixes with the rising solvent. The dephenolized wastewater is stripped with steam to recover residual solvents before it is sent to subsequent treatment. The rich solvent is sent to a solvent recovery column where the solvent is distilled off as overhead and the phenols are removed as bottoms from the column.

The performance of the Chem-Pro process is dependent on a number of factors such as the characteristics of the waste stream, the solvent-to-waste ratio and the column design. Approximate maximum removals of phenols from feed waters are estimated by the process developer as: 99.8% removal of monohydric phenol, 95% removal of dihydric phenol, 90 to 95% removal of trihydric phenol and 50 ± % removal of other organics, depending on the characteristics of the organics (60,66). The breakdown of the phenols in an EDS combined wastewater is about 90% monohydric and 10% dihydric (62). Thus it is expected that about 99% of the total phenols will be removed by this process (see Table 4-49). There is no data on the bulk organic removal efficiencies of this process. Since the process principle is similar to the Phenosolvan process, the TOC and COD removal efficiencies are assumed to be the same for the two processes.

The estimated capital and annualized costs for this process are summarized in Table 4-50. Using the procedures summarized in Table 4-1 the capital investment is estimated based upon the installed equipment cost reported in the literature (60). The capital investment for the two cases are 12.4 and 10.8 million dollars, or 0.39% and 0.37% of the base plant cost. Using the procedures summarized in Table 4-2 and assuming the utility and other requirements are the same as those for the Phenosolvan process, the total annualized costs for the base and MFS case are estimated to be 2.6 and 2.2 million dollars or about $\$1.3/m^3$ and $\$1.4/m^3$, respectively. As discussed in Section 4.3.1.1.2.1, when estimating the annualized cost, recovered phenol is credited with its heating values only (2.2¢/kg), which is much lower than its potential resale value (20¢/kg). If the phenol were to sell for 20¢/kg, the base case and the MFS case will generate a profit of 0.75 and 0.70 million dollars per year, respectively.

Secondary Streams

The Chem-Pro process generates two waste streams: the treated effluent and the recovered phenol. The estimated characteristics of the treated effluent are presented in Table 4-49. It is not possible to estimate the characteristics of the recovered phenol because of the potential presence of other organics. Nonetheless, based on the removal efficiency, it is expected that the relative concentrations of monohydric phenol/polyhydric phenol/organic acids in the combined stream for the base case and MFS case will be 77%/8%/15% and 76%/8%/16%, respectively.

TABLE 4-49. ESTIMATED COMPOSITION OF FEED AND TREATED EFFLUENT FOR CHEM-PRO PROCESS (STREAM A) (1,134 Mg/hr Coal Feed EDS Plant)

	Base Case		MFS Case	
Component	Combined Stream Feed	Treated Effluent	Combined Stream Feed	Treated Effluent
Phenols, mg/l	9,300	93	11,000	110
TOC, mg/l	14,000	4,200	16,000	4,800
COD, mg/l	45,000	18,000	51,000	20,000
Organic Acids, mg/l	3,200	1,600	4,000	2,000
Flow, m^3/hr	260	260	210	210

TABLE 4-50. SUMMARY OF COST ESTIMATES FOR THE CHEM-PRO PROCESS (STREAM A)

Cost Item	Base Case	MFS Case
Capital Investment	$\$12.4 \times 10^6$	$\$10.8 \times 10^6$
Total Annualized Cost	$\$2.6 \times 10^6$	$\$2.2 \times 10^6$
% Uncontrolled Base Plant Capital Cost	0.39	0.37
% Uncontrolled Plant Annualized Cost	0.27	0.24
Unit Capital Investment	$48,000/$m^3$/hr	$51,000/$m^3$/hr
Unit Annualized Cost	$1.3/$m^3$	$1.4/$m^3$

4.3.1.1.2.3 Control Technique 3 - Resin Adsorption

While the solvent extraction technique removes the phenols from wastewater with an organic solvent, resin adsorption techniques remove the phenol by a solid medium (the resins). Resin adsorption is a cyclic process where phenol is recovered from wastewater. The cyclic process consists of three steps: adsorption on the resin, regeneration followed by a water rinse, and separation of the regenerant (acetone) and the phenolic compounds by distillation. Adsorption is carried out in resin beds where wastewater enters through the top and exits through the bottom of the bed. Regeneration is achieved by rinsing the bed with acetone (67).

The performance of the resin adsorption process is dependent on a number of factors such as the characteristics of the waste stream, the bed loading rate and bed height, and frequency of regeneration. It has been reported that less than 10 ppm phenols in the treated effluent has been achieved when treating various industrial wastewaters with initial phenol concentration ranging from a few hundred ppm to 10,000 ppm (67). There is no data on removal efficiencies for organics other than phenolics.

The estimated capital and annualized cost for this process are summarized in Table 4-51. The capital investment ranges from 8.7 to 9.9 million dollars or about 0.31% of the base plant cost. The total annualized costs for the base and MFS case are estimated to be 4.4 and 3.6 million dollars, or about $\$2.1/\mathrm{m}^3$ and $\$2.2/\mathrm{m}^3$, respectively. The capital cost is factored from reported literature value on a plant treating wastewater from a Bisphenol A manufacturing plant (64). The high annualized cost for this process is due to the high energy consumption during regeneration and distillation steps; (this is estimated to be 250 Mg of steam/m^3 of wastewater treated). As discussed in Section 4.3.1.1.2.1, when estimating the annualized cost, recovered phenol is credited with its heating value only (2.2¢/kg), which is much lower than its potential resale value (20¢/kg). If the phenols were to sell for 20¢/kg, the annualized costs for the base case and MFS case would be $990,000 and $530,000, respectively.

TABLE 4-51. SUMMARY OF COST ESTIMATES FOR THE RESIN ADSORPTION PROCESS - STREAM A (1,134 Mg/hr Coal Feed EDS Plant)

Cost Item	Base Case	MFS Case
Capital Investment	$\$9.9 \times 10^{6}$	$\$8.7 \times 10^{6}$
Total Annualized Cost	$\$4.4 \times 10^{6}$	$\$3.6 \times 10^{6}$
% Uncontrolled Base Plant Capital Cost	0.31	0.31
% Uncontrolled Base Plant Annualized Cost	0.46	0.41
Unit Capital Investment	$\$38{,}000/\text{m}^3/\text{hr}$	$\$41{,}000/\text{m}^3/\text{hr}$
Unit Annualized Cost	$\$2.1/\text{m}^3$	$\$2.2/\text{m}^3$

Note: Capital investment is estimated using Table 4-1 and installed equipment cost from Reference 67. Annualized cost is estimated using Table 4-2 and the following utilities/chemicals requirements derived from Reference 67: electricity - 0.87 kwh/m^3, stream - 0.25 Mg/m^3, acetone - $0.43/\text{m}^3$ and resin - $\$0.25/\text{m}^3$ of feed.

Secondary Streams

The resin adsorption generates two secondary waste streams: the treated effluent and recovered phenol. There is no data for either of these streams. The treated effluent is expected to contain less than 10 ppm phenol.

4.3.1.1.2.4 Control Technique 4: Wet Air Oxidation

The wet air oxidation process works on the principle that under high pressures and elevated temperatures reduced organic and inorganic substances can be oxidized while remaining in the liquid stage. In this process wastewater is pumped through a preheater where its temperature is raised to about 500°K (680°F) before entering the reactor. The process is carried out under pressure (up to 208 MPa; 3000 psig) in order to prevent boiling and to increase oxygen solubility. Compressed air or oxygen is added to the wastewater after preheating.

The degree of oxidation achieved depends upon the retention time in the reactor, the reactor temperature and pressure, and the material being oxidized. When organic COD exceeds 15,000 mg/l the process is autothermic and requires no additional energy input. COD removals of greater than 90% have been reported for this process (68). This process will oxidize both organics and inorganic COD, including conversion of sulfides to sulfate. Ammonia will pass through unaffected. Table 4-52 summarizes the expected performance of this process when applied to the EDS combined waste stream (see Appendix B-6 for details). One major advantage of this process is that complex organic molecules which are not normally biodegradable will be destroyed, making the effluent more amenable to biotreatment; this may also reduce the costs of biological treatment if it is required. On the other hand, being an oxidation process, the phenols in the EDS wastewater will also be destroyed and thus cannot be recovered as salable by-product.

WAO requires high capital investment. Special materials of construction (such as Hastelloy or titanium) are required for the reactor and high temperature exchangers because of the corrosive process environment. Based on the

TABLE 4-52. ESTIMATED COMPOSITION OF FEED AND TREATED EFFLUENT FROM THE WET AIR OXIDATION PROCESS - STREAM A (1,134 Mg/hr Coal Feed EDS Plant)

Component	Base Case Combined Stream Feed	Base Case Treated Effluent	MFS Case Combined Stream Feed	MFS Case Treated Effluent
Phenols, mg/l	9,300	93	11,000	110
TOC, mg/l	14,000	1,400	16,000	1,600
COD, mg/l	45,000	2,200	51,000	2,500
Organic Acid, mg/l	3,210	640	4,000	800
H_2S, mg/l	18,000	1,800	22,000	2,200
Flow, m^3/hr	260		210	

Note: See Appendix B-6 for bases and assumptions.

process vendor's estimates, the capital investments will be 72 and 63 million dollars or \$280,000/m^3/hr and \$300,000/m^3/hr for the base case and MFS case, respectively (69).

The wet air oxidation process produces no secondary waste streams other than the treated effluent. The characteristics of this stream are presented in Table 4-52.

4.3.1.1.3 Control Function 3 - Removal of Dissolved Gases

The Combined Sour Water Stream contains high levels of ammonia, H_2S and CO_2. Removal of these gases at such high concentration ranges is normally accomplished by steam stripping. In all existing or proposed coal conversion facilities, stripping is aimed also at the recovery of ammonia for sale as a by-product. This section evaluates two such processes: the Phosam-W Process and the Chevron WWT Process. As discussed before, these techniques can be placed ahead of or after Control Function 2 techniques.

4.3.1.1.3.1 Control Technique 1 - Phosam-W Process

In the Phosam-W Process, wastewater enters a steam stripper where free NH_3, H_2S, CO_2, HCN and volatile organics are stripped. The stripper overhead gases then enter an ammonia absorber where lean ammonium phosphate solution contacts the sour vapor, absorbing ammonia and small amounts of other gases. Stripped wastewater leaves the bottom of the stripper. Rich ammonium phosphate solution is purged of acid gases in a contactor and sent to a stripper for ammonia removal. The ammonia recovered is separated from water by distillation. The other purged acid gases usually go to the sulfur recovery plant.

The process has been reported to be capable of removing NH_3 to below 200 ppm, and essentially all the H_2S and CO_2 (70,71). Table 4-53 presents the characteristics of the influent and effluent streams, and also of the by-product ammonia and the purged gases. The ammonia by-product stream, with a

TABLE 4-53. ESTIMATED COMPOSITION OF FEED AND TREATED EFFLUENT STREAMS FOR THE PHOSAM-W PROCESS (1,134 Mg/hr Coal Feed EDS Plant)

Stream	Component, kg/hr (mg/l) H_2O	H_2S	CO_2	NH_3	Total Flow (m^3/hr)
Base Case					
Combined Feed	250,000	4,700	2,300	3,600	260
Stripper Overhead	16,000	4,700	2,300	nil	--
Recovered Ammonia	nil	nil	nil	3,600	--
Treated Effluent	230,000	nil	nil	50(200)	230
MFS Case					
Combined Feed	200,000	4,600	2,100	3,500	210
Stripper Overhead	16,000	4,600	2,100	nil	--
Recovered Ammonia	nil	nil	nil	3,500	--
Treated Effluent	180,000	nil	nil	40(200)	180

TABLE 4-54. SUMMARY OF COST ESTIMATES FOR THE PHOSAM-W* PROCESS

Cost Item	Base Case	MFS Case
Capital Investment	$\$10.8 \times 10^6$	$\$9.52 \times 10^6$
Total Annualized Cost (profit)	(1.56×10^6)	(1.05×10^6)
% Base Plant Capital Cost	0.34	0.33
% Base Plant Annualized Cost	-	-
Unit Capital Investment	\$42,000/m^3/hr	\$45,000/m^3/hr
Unit Annualized Cost (profit)	(\$0.80/m^3)	(\$0.60/m^3)

* Capital investment is estimated using the procedures summarized in Table 4-1 and the unit installed equipment cost presented in Figure B8-2 (Appendix B-8). Annualized cost is obtained using bases presented in Table B8-2 and the unit cost in Table 4-2.

purity of 99.99% , can be sold as a product. The acid gas stream which consists mainly of H_2S and CO_2, is usually routed to the sulfur recovery plant for sulfur recovery. The stripped wastewater, which contains about 200 ppm NH_3 and other contaminants, may require further treatment before disposal or reuse.

Table 4-54 presents the capital and annualized costs for the Phosam-W process when applied to the EDS Combined Stream. The capital investment is factored from reported literature values (see Figure B-8-2 of Appendix B-8). The detailed basis and assumptions for the cost estimates are presented in Appendix B-8. As shown in Table 4-54 the capital investments are $10.8 and $9.5 million for the base and MFS case, respectively. Because of the high resale value of the ammonia recovered, which is assumed to be $140/Mg, this process generates a net profit of $1.6 and $1.1 million/year for the two cases. If there were no market for the ammonia, the annual cost would be $2.5 and $2.9 million per year, respectively, for the base case and MFS case.

Secondary Streams

The Phosam-W process generates three by-product/waste streams, namely an ammonia by-product stream, an acid gas stream, and a stripped wastewater stream. The characteristics of these streams have been presented in Table 4-53.

4.3.1.1.3.2 Control Technique 2 - Chevron WWT Process

In the Chevron WWT Process, hydrogen sulfide and ammonia in the sour waters are stripped in two separate stages to produce gaseous streams suitable for sulfur and ammonia recovery. Feed wastewater is fed to a reboiler stripper column (H_2S stripper) where hydrogen sulfide and carbon dioxide are stripped. Stripper bottoms with the bulk of the ammonia are fed to a second reboiler stripper column (NH_3 stripper) for ammonia stripping. The H_2S stripper is operated at 690 KPa (100 psig) and 367°K (200°F), and the NH_3 stripper is operated at 345 KPa (50 psig) and 394°K (250°F) (71). Because H_2S is more insoluble in water than NH_3, heating the H_2S stripper to about 367°K (200°F) causes the H_2S but not the NH_3 to be released in that column. The overhead from the NH_3 stripper is scrubbed with cold aqueous ammonia to remove traces of H_2S and is compressed and condensed to form anhydrous ammonia.

It is reported that the process is capable of producing a hydrogen sulfide stream with less than 50 ppm NH_3, and less than 5000 ppm H_2O; an ammonia stream with less than 5 ppm H_2S; and a stripped water stream containing less than 50 ppm ammonia and 5 ppm sulfide (72,73). Table 4-55 summarizes the characteristics of these streams. The ammonia stream, with less than 5 ppm H_2S and 1000 ppm H_2O, can be sold as by-product. The acid gas stream which consists mainly of H_2S and CO_2, is usually routed to the sulfur recovery plant for sulfur recovery. The stripped wastewater, which contains 50 ppm of NH_3 and other contaminants, may require further treatment before disposal or reuse.

Table 4-56 summarizes the capital and annualized costs for the Chevron-WWT process when applied to the EDS Combined Stream. The capital investment is factored from reported literature values (see Figure B-9-2 of Appendix B-9). The detailed bases and assumptions for the cost estimates are presented in Appendix B-9. As shown in Table 4-56 the capital investments are \$11.8 and \$10.7 million, respectively, for the base case and MFS case. The total annualized costs are \$2.2 and \$1.4 million. This cost includes a credit of \$14/Mg for the ammonia recovered. If there were no market for the ammonia, the annualized cost would be increased to \$6.2 and \$5.2 million per year.

Secondary Streams

The Chevron WWT process generates three by-product/waste streams, namely, an ammonia by-product stream, an acid gas stream (from the H_2S stripper) and a stripped wastewater stream. The characteristics of these streams are presented in Table 4-55.

TABLE 4-55. ESTIMATED COMPOSITION OF FEED AND TREATED EFFLUENT STREAMS FOR THE CHEVRON-WWT PROCESS (1,134 Mg/hr Coal Feed EDS Plant)

Stream	Component, Kg/hr (mg/l)				Total Flow (m^3/hr)
	H_2O	H_2S	CO_2	NH_3	
Base Case					
Combined Feed	250,000	4,700	2,300	3,600	260
Stripper Overhead	30 (5,000)	4,	2,	<1 (50)	--
Recovered Ammonia	4 (1,000)	<1 (5)	nil	3,600	--
Treated Effluent	200,000	1 (5)	nil	10 (50)	250
MFS Case					
Combined Feed	200,000	4,600	2,100	3,500	210
Stripper Overhead	30 (5,000)	4,600	2,100	<1 (50)	--
Recovered Ammonia	4 (1,000)	<1 (5)	nil	3,500	--
Treated Effluent	200,000	1 (5)	nil	10 (50)	200

TABLE 4-56. SUMMARY OF COST ESTIMATES FOR THE CHEVRON-WWT PROCESS*

Cost Item	Base Case	MFS Case
Capital Investment	$\$11.8 \times 10^6$	$\$10.7 \times 10^6$
Total Annualized Cost	$\$2.2 \times 10^6$	$\$1.4 \times 10^6$
% Base Plant Capital Cost	0.37	0.37
% Base Plant Annualized Cost	0.23	0.15
Unit Capital Investment	\$45,000/m^3/hr	\$51,000/m^3/hr
Unit Annualized Cost	\$1.1/m^3	\$0.84/m^3

* Capital investment is estimated using the procedures summarized in Table 4-1 and the unit installed equipment cost presented in Figure B9-2 (Appendix B-9). Annualized cost is obtained using the bases presented in Table B9-3 and the unit cost presented in Table 4-2.

4.3.1.1.4 Control Function 4 - Removal of Dissolved Organics

Table 4-57 summarizes the estimated characteristics of the Combined Stream A after Function 2 and 3 controls. The ranges for the various species shown result from the ranges of performances by the different alternative Function 2 and 3 techniques. As shown, the stream still contains high levels of organics. Common practice to remove dissolved organics is by some form of biological treatment. In simple terms, biological treatment is a process which provides an environment conducive to microorganism reduction of organic materials in the wastewater. There are many versions of biological treatment processes that are commercially available. These include trickling filters, rotating biodiscs, oxidation ponds/lagoons, and various modifications of activated sludge. Of all these processes, activated sludge is the most studied for use in treating coal conversion wastewaters and is the most commonly used in existing industrial and municipal applications. There is generally more information on this process than other processes. In actual application, an optimum biological system could be a combination of these processes. For example, it has been proposed that using a trickling filter ahead of an activated sludge system would reduce the overall control cost (74). However, design and performance data for these processes are generally lacking. For these reasons, activated sludge is the only process evaluated in this section as an example of biological treatment processes.

4.3.1.1.4.1 Control Technique 1 - Activated Sludge Process

All activated sludge processes employ one or more reactors in which aerobic microorganisms feed on the organic matter present in the wastewater. These microorganisms form colonies which ultimately appear as a sludge which must be separated from the wastewater. A portion of the settled sludge is returned to the reactor as needed to insure a sufficient population of microorganisms in the reactor. Air or oxygen is mixed into the reactor to provide an adequate supply of oxygen for the overall oxidation process.

TABLE 4-57. ESTIMATED CHARACTERISTICS OF COMBINED STREAM A AS FEED TO CONTROL FUNCTION 4 (1,134 Mg/hr Coal Feed EDS Plant)

Component	Estimated Concentration, mg/l*	
	Base Case	MFS Case
H_2S	<5	<5
NH_3	50 - 200	50 - 200
CO_2	nil	nil
Phenols	93 - 460	110 - 550
Organic Acid	640 - 2,700	800 - 3,400
COD	2,200 -18,000	2,500 -20,000
TOC	1,400 - 4,200	1,600 - 4,800
Flow, m^3/hr	260	210

*After Functions 2 and 3 controls.

Biological treatment of coal conversion process wastewaters has not been demonstrated on a commercial basis. Bench scale studies by several investigators have proven the feasibility of air activated sludge systems when applied to wastewaters from several gasification and liquefaction processes (75-82). COD levels in these studies run as high as 12,000 mg/l. COD feed to commercial scale activated sludge plants treating domestic or industrial wastes generally range from several hundred mg/l to several thousand mg/l. In an actual EDS wastewater treatment plant, dilution of the combined stream may be required. This can be accomplished by recycling a portion of the treated effluents or combining with other wastewaters such as cooling tower blowdowns. In addition, the activated sludge system will likely be a multi-staged system, and high purity oxygen or activated powdered carbon will likely be used to enhance the stability of the system. Nutrients such as phosphorus will need to be added.

Table 4-58 summarizes the anticipated performance of this process. Since no EDS wastewater treatability data are publicly available, 80% removal for both the COD and TOC contents were assumed. This is consistent with several bench scale studies treating different coal conversion process wastewaters (75-82). See Appendix B-10 for more details.

The ammonia concentration in the treated effluent is dependent on the ammonia level in the feed to the biox system and the degree of nitrification (conversion of ammonia to nitrite/nitrate) in the reactor. The optimum organic to nitrogen ratio in the feed to biox systems is about 20:1. As shown in Table 4-58, this ratio is much higher in the EDS streams, i.e., nitrogen is the limiting nutrient and it must be added to the feed. The level of ammonia in the feed to the biotreatment unit may be adjusted by changing the design of Control Function 3 technique (ammonia recovery techniques), or by adding compounds such as ammonium nitrate and ammonium sulfate. The effluent ammonia concentration would thus depend on how much 'excess' ammonia is added. It is expected that much of the 'excess' ammonia will be nitrified since long

TABLE 4-58. ESTIMATED COMPOSITION OF FEED AND TREATED EFFLUENT FOR THE ACTIVATED SLUDGE PROCESS - STREAM A (1,134 Mg/hr Coal Feed EDS Plant)

Component	Base Case		MFS Case	
	Combined Stream Feed	Treated Effluent	Combined Stream Feed	Treated Effluent
COD, mg/l	2,200-18,000	440-3,600	2,500-20,000	500-4,000
TOC, mg/l	1,400- 4,200	280 - 840	1,600- 4,800	320 - 960
Phenol, mg/l	93 - 460	5 - 23	110 - 550	6 - 28
Organic Acid, mg/l	640-2,700	32 - 140	800- 3,400	40 - 170
NH_3, mg/l	50 - 200	10	50 - 200	10
Flow, m^3/hr	260	260	210	210

TABLE 4-59. SUMMARY OF COST ESTIMATES FOR THE ACTIVATED SLUDGE PROCESS*

Cost Item	Base Case	MFS Case
Capital Investment	\$20 x 10^6	\$16 x 10^6
Total Annualized Cost	\$ 4.9 x 10^6	\$ 4.2 x 10^6
% Uncontrolled Base Plant Capital Cost	0.63%	0.56%
% Uncontrolled Base Plant Annualized Cost	0.52%	0.47%
Unit Capital Investment	\$77,000/m^3/hr	\$ 76,000/m^3/hr
Unit Annualized Cost	\$ 2.4/m^3	\$ 2.5/m^3

*Capital investment is estimated from top curve in Figure B10-2 (Appendix B-10), using BOD loading values of 79,000 Kg/day and 71,000 Kg/day, respectively, for the base and MFS case (BOD was assumed to be 0.7 x COD). Annualized cost is derived from Figure B10-3 and assumptions presented in Table 4-2.

residence time in the bioreactor will be required to handle the organics. However, some residual ammonia will likely be present in the effluent. For this study, this is assumed to be 10 mg/l.

The capital investment and total annualized costs for the two cases are summarized in Table 4-59. The values presented are the costs required to treat the higher concentration range of organic (i.e., the worst case) as shown in Table 4-58. The capital investment for the two cases are 20 and 16 million dollars, or about 0.63% and 0.56% of the base plant cost. The total annualized costs are 4.9 and 4.2 million dollars per year, or about \$2.4 and \$2.5/m^3 of treated water.

Secondary Streams

The activated sludge process generates a secondary waste stream, namely, the waste sludge. This stream is generally stabilized by anaerobic digestion, dewatered to 20 to 40% solid and then disposed. This stream may also be stabilized by aerobic digestion and incineration. The amount of dewatered sludges generated in the base case and MFS case are estimated to be, respectively, 100 and 80 metric ton per day. The sludge is assumed to contain 80% moisture and 20% solid. The bulk of the solids are bacterial cell walls. Depending on the characteristics of the feed to the biox system, some toxic nonbiodegradable organics such as PAH's may also be accumulated in the sludge through sorption mechanisms.

4.3.1.1.5 Control Function 5 - Removal of Residual Organics

Available data on several coal conversion process wastewaters indicate high ratios of COD/BOD in the raw wastewater. This is believed to be caused by the presence of high molecular weight, complex, refractory organics. Although there is no data on the full characterization of Combined Stream A, it is likely that the stream will contain many nonbiodegradable organic molecules. These molecules will remain in the stream unless specific treatment processes are employed for their removal. This section evaluates three techniques for removing these organics. Feed wastewater to these techniques will generally be treated effluent from biox system (see Table 4-58).

4.3.1.1.5.1 Control Technique 1 - Activated Carbon Adsorption

Activated carbon adsorption is one of the most efficient and commercially proven methods of removing organics from industrial wastewaters. Primarily for economic reasons, it is generally used as a "polishing" step after the majority of the organics are removed by other processes. The application of activated carbon may be carried out by two methods: (1) the use of powdered activated carbon in conjunction with a biological treatment or other mixed unit and (2) the direct contact of granular activated carbon (GAC) with contaminated wastewaters following biox. The direct contact method will be evaluated in the remainder of this section because this is the most widely used approach and because most of the available performance data on activated carbon systems are for this application.

Removal of organics with granular activated carbon is accomplished by passing the wastewater through a packed bed of the adsorbent. The adsorptive capacity of the bed is a function of the wastewater's organic loading and contact time in the bed, and the affinity of the carbon for the organics. Different types of activated carbon can generally be selected to obtain high percentage removals of various organic compounds. Carbon can act as both an adsorption medium for smaller organic molecules and a filter for long chain, high molecular weight molecules. Carbon removal efficiency is not sensitive to changes in organic loading. But, organic concentrations above design loadings can significantly increase regeneration frequency and the energy requirements for operating the system.

The characteristics of the treated effluent stream is presented in Table 4-60. Also presented in the table are the characteristics of the feed to the activated carbon system. This feed is assumed to be the Combined Sour Stream after going through Function 4 control. Since detailed organic characteristics of the feed is not known and no treatability data are available, the table is constructed based on available literature data on other related waste streams (83-86). It should be emphasized that the removal efficiency is very sensitive to the characteristics of the waste stream. Actual performance on Combined Stream A can only be established by treatability studies with the actual stream.

The capital investment and annualized cost for this process are summarized in Table 4-61. The values presented are the costs required to treat the higher concentration range of organics as shown in Table 4-60, i.e., the worst case. The capital cost, which includes the thermal regeneration system, are 2.7 million dollars for both cases. The total annualized costs are 1.5 and 1.4 million dollars for the base case and MFS case, respectively; or about $\$0.73/m^3$ and $\$0.93/m^3$ of treated water.

Secondary Streams

Activated carbon adsorption system requires regeneration when the adsorptive capacity of the bed is reached. Thermal regeneration is the most common system used. Most of the organics adsorbed by the carbon are expected to be destroyed during regeneration. During regeneration an off-gas stream is generated. This stream is expected to contain particulates, CO and other constituents, depending on the organics adsorbed. It is common practice that scrubbers are included as part of the regeneration system. Thus the particulate contents in the off-gas leaving the regeneration is expected to be minimal. The scrubber wastewater is typically settled or routed to other treatment before disposal.

TABLE 4-60. ESTIMATED COMPOSITION OF FEED AND TREATED EFFLUENT FOR THE ACTIVATED CARBON ADSORPTION PROCESS - STREAM A
(1,134 Mg/hr Coal Feed EDS Plant)

	Base Case		MFS Case	
Component	Influent	Effluent	Influent	Effluent
COD, mg/l	440 - 3,600	44 - 360	500 - 4,000	50 - 400
TOC, mg/l	280 - 840	28 - 84	320 - 960	32 - 96
Phenol, mg/l	5 - 23	0.05-2	6 - 28	0.06-3
Organic Acid, mg/l	32 - 140	10 - 40	40 - 170	12 - 50
Flow rate, m^3/hr	260	260	210	210

TABLE 4-61. SUMMARY OF COST ESTIMATES FOR THE ACTIVATED CARBON ADSORPTION PROCESS*

Cost Item	Base Case	MFS Case
Capital Investment	$\$ 2.7 \times 10^6$	$\$ 2.7 \times 10^6$
Total Annualized Cost	$\$ 1.5 \times 10^6$	$\$ 1.4 \times 10^6$
% Base Plant Capital Cost	0.09%	0.09%
% Base Plant Annualized Cost	0.16%	0.15%
Unit Capital Investment	\$10,000/m^3/hr	\$13,000/m^3/hr
Unit Annualized Cost	\$ 0.73/m^3	\$ 0.86/m^3

* Capital investment is obtained by using the procedure summarized in Table 4-1 and summing the installed equipment cost for absorption contactor (Figure B15-1, Appendix B-15) and regeneration furnace (Figure B15-3), contactor volumes of 130 m^3 and 105 m^3 were used for the base case and MFS case, respectively. These volumes were derived assuming an empty bed contact time of 30 minutes. Carbon regeneration rates for the two cases were determined assuming a carbon adsorption capacity of 0.5 Kg COD/Kg carbon. Annualized cost was determined using the utility and other requirements summarized in Appendix B-15 and the unit costs presented in Table 4-2.

4.3.1.1.5.2 Control Technique 2 - Chemical Oxidation

Chemical oxidation, as its name implies, is simply the oxidation of wastewater contaminants, both organic and inorganic, to more environmentally acceptable forms. In practice, the wastewater is introduced to one or several reactors where chemical oxidants are added. Common chemicals used included chlorine, ozone, hydrogen peroxide, potassium dioxide and chlorine dioxide. Some recently developed processes employ ultraviolet (UV) light with one other strong oxidant such as ozone, hydrogen perioxide or chlorine. The UV light would excite the organic molecules, making them more susceptible to the chemical oxidant attack. Due primarily to economic reasons, chemical oxidation process has been limited to use as a polishing step, treating low concentration waste streams.

The performance of chemical oxidation is dependent on a number of factors such as the waste stream characteristics, the chemical and the dosage used, and reactor design. Greater than 90% removal in various types of industrial wastewater have been reported (87-89). However, no treatability test data on the EDS wastewater are publicly available. Nevertheless, in principle, any degree of removal can be achieved if enough chemicals and detention time are provided, and the waste is oxidizable. However, treating the subject stream to any significant degree could incur high costs. As presented in Table 4-58, the COD of feed to this system may be as high as 3,600 ppm and 4,000 ppm for base case and MFS case, respectively. Theoretically, 3 kg of ozone is required to remove 1 kg of COD. Applying ozone as treatment to this stream would incur capital investments of 97 and 87 million dollars for the two cases, or about \$370,000/m^3/hr and \$410,000/m^3/hr; the annualized costs would be \$42 and \$38 million dollars, or about \$20/m^3 and \$23/m^3. These costs are developed using the assumptions presented in Tables 4-1 and 4-2, and the installation costs, electricity requirement and oxygen requirement estimates presented in reference 89. If chlorine is used, theoretically, 4.5 parts chlorine are required for each part of COD removed. Assuming a chlorine cost of $180/Mg (89), just the annual chlorine costs alone would amount to 6 and 5 million dollars for the two cases. Chemical oxidation is further discussed in Appendix B-16.

4.3.1.1.5.3 Control Technique 3 - Incineration

Incineration involves oxidizing the organics at high temperature. Most organics can be destroyed at 1000°C with two seconds residence time. In practice, wastewater is injected into the incinerator through a liquid injection system where the organics are destroyed. The flue gas leaving the incinerator is cooled and quenched before discharge to the atmosphere.

Incineration is generally economical only for treating low volume, high strength wastes because of the costs of auxiliary fuel and the equipment required in order to heat and vaporize the entire mass of wastewater (90,91). Approximately 2500 Kcal of energy is required to incinerate one kg of wastewater (92). Thus, when incineration is applied to removing residual organics in the Combined Sour Water Stream, some form of volume reduction prior to incineration may be required. Table 4-62 presents the estimated feed and treated effluent from the incinerator. It is assumed that the feed is the treated effluent from Function 4 controls, and have been concentrated ten times. The concentration technique is discussed in the next section (Function 6). The effluent composition (after re-condensation) corresponds to 99.9 percent destruction of thermally oxidizable constituents. As discussed in Appendix B-17 incineration is capable of greater than 99.96% destruction of many organic compounds. The 99.9% destruction efficiency shown in Table 4-62 is therefore felt to be a conservative estimate.

The capital investment and the annualized cost for this process are summarized in Table 4-63. The detailed basis and assumption for the cost estimates are presented in Appendix B-17. The capital investment, which includes the quench and scrubber system, are 5 and 4.6 million dollars for the base case and MFS case or about 0.16% of the base plant cost, respectively. The

TABLE 4-62. ESTIMATED COMPOSITION OF FEED AND TREATED EFFLUENT FOR WASTEWATER INCINERATION - STREAM A (1,134 Mg/hr Coal Feed EDS Plant)

Component	Base Case		MFS Case	
	Combined Stream Feed*	Treated Effluent	Combined Stream Feed*	Treated Effluent
NH_3, ppm	100	0.1	100	0.1
Phenol, ppm	230	0.23	280	0.28
TOC, ppm	8,400	8.4	9,600	9.6
COD, ppm	36,000	36	40,000	40
Organic Acid, ppm	1,400	1.4	1,700	1.7
TDS	40,000	40,000	49,000	49,000
Flow Rate, M^3/hr	26	26	21	21

*Assumed to be 10-fold concentration of treated effluent from function 4.

TABLE 4-63. SUMMARY OF COST ESTIMATES* FOR THE WASTEWATER INCINERATION

Cost Item	Base Case	MFS Case
Capital Investment	$\$5.0 \times 10^6$	$\$4.6 \times 10^6$
Total Annualized Cost	$\$5.2 \times 10^6$	$\$4.3 \times 10^6$
% Base Plant Capital Cost	0.16%	0.16%
% Base Plant Annualized Cost	0.55%	0.48%
Unit Capital Investment	\$190,000/m^3/hr	\$220,000/m^3/hr
Unit Annualized Cost	\$25/m^3	\$26/m^3

* Capital investment is estimated using the procedures summarized in Table 4-1 and the installed equipment cost presented in Figure B17-2 (Appendix B-17). Incinerator duties of 650×10^6 Kcal/hr and 530×10^6 Kcal/hr are assumed for the base case and MFS case, respectively. Annualized cost is estimated by using the bases presented in Table 4-2, and the following utility and labor requirements: 2500 Kcal of fuel/Kg of water and 2 person/shift.

total annualized costs are 5.2 and 4.3 million dollars, or about \$25/m^3 to \$26/m^3 of treated water.

Secondary Streams

The only secondary waste stream is the incinerator flue gas. This stream may contain SO_2, NO_x and particulates. The major contribution to the SO_2 and NO_x is likely from the fuel used. The particulates may come from the fuel and from dissolved solids in the wastewater. Since control for this stream is considered as part of the incinerator system*, no estimates on the characteristics or cost are presented for this stream.

4.3.1.1.6 Control Function 6 - Volume Reduction

Volume reduction is one form or part of a water reuse scheme. Water reuse is generally practiced in arid regions where water is scarce, or regions where surface discharge requirements are very stringent (such as no discharge) and other disposal capacities (such as underground injection) are limited. This function involves concentrating the pollutants into smaller waste streams for further treatment or disposal; water of different qualities may be recovered for reuse in the process, depending on the techniques used and the feed characteristics. To prevent excessive air emission of volatile contaminants and to ensure the water recovered is of reasonable quality, this function is likely used after or in lieu of Function 5.

4.3.1.1.6.1 Control Technique 1 - Cooling Tower

Based upon the availability of cooling water at the plant site, the use of process wastewater as partial or total feed to the cooling tower may be practiced. This practice results in a reduction in the waste stream volume due to evaporative concentration in, and drift from, the cooling tower. Cooling towers typically run in 4 to 5 cycles of concentration, representing a 75 to 80% volume reduction, i.e., 75 to 80% of the water is evaporated.

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* Of the 83 installed incinerators designed to burn aqueous waste containing sodium salts, 80 have a quench tank followed by a venturi scrubber, 3 have electrostatic precipitators (91).

It is likely that the Combined Sour Water Stream would at least undergo Functions 2, 3 and 4 controls and probably Function 5 control, before being used as cooling tower makeup. Using the untreated stream would generate excessive air emissions from the cooling tower of volatiles such as phenol, H_2S and NH_3 contained in the stream. Drift could contain even the non-volatile species. Also, although cooling tower is conducive to organics reduction due to biological activities, passing the non-biologically treated stream through the cooling tower would also increase potential for fouling process heat exchangers by wastewater contaminants such as biological solids, and by excessive biogrowth. To overcome this problem, redundant heat exchangers would possibly be required to maintain operating reliability. These redundant heat exchangers would be designed in such a way so that individual heat exchangers could be isolated and backflushed while the back-up system is on-line (93). The redundant heat exchanger design and the isolation and capability which has been used in one refinery operation (93) would greatly increase the EDS base plant cost and likely offset any cost savings for using the untreated stream as makeup.

The cooling operation is an integral part of the process waste heat rejection system and is thus part of the base plant. It was assumed that no additional costs to the plant are incurred as a result of using the cooling tower as a wastewater concentrator.

Secondary Stream

Cooling tower operation generates three streams: evaporation, drift and blowdown. Assuming all the volatile components of the waste stream have been removed by Function 2, 3, 4, and possibly 5 controls, water evaporated from the cooling tower is expected to be relatively clean, but could contain traces of the non-volatile species. The drift and the blowdown are expected to be similar in composition and would depend on the make-up characteristics and cycles of concentrations the cooling tower is operating at. To improve the quality of these two streams, Function 5 techniques can be used ahead of the cooling tower. Table 4-64 presents the estimated characteristics of these streams. The feed ranges shown in Table 4-64 cover the ranges of treated effluents from Function 4 (in the event that 5 is not employed) and from Function 5. The cooling tower is assumed to be using exclusively the treated wastewater as make-up, and is operated at 4 cycles of concentration. The drift and the blowdown are assumed, respectively to be 0.03% and 3% of the circulating water. These are typical design/operating numbers for cooling towers (92).

4.3.1.1.6.2 Control Technique 2 - Vapor Compression Evaporator

Vapor compression evaporator is used as an example of reducing the volume of wastewater by forced evaporation. Other commercially available force evaporators include multi-effect and multi-stage evaporators. Although costs and performances vary among these different designs, for the purpose of this manual, they would probably fall into the same ranges.

The vapor compression evaporator supplies vaporization energy by a mechanical compressor, rather than an external heating medium. The wastewater feed is first treated in a feed tank to adjust the pH to carbon dioxide for removal in a deaerator. The waste stream is then pumped through a feed-product heat exchanger to raise its temperature to about its atmospheric boiling point. The heated feed then enters a counterflow packed column deaerator which strips carbon dioxide, nitrogen, and oxygen.

Next, the feed enters the evaporator sump and is mixed with concentrated slurry. The brine slurry is constantly circulated from the sump to the top of the evaporator tubes. At the top of the vertical-tube, falling film evaporator, the brine slurry is distributed to the inside wall of the tubes. The slurry flows as a film on the inside of the tubes down to the sump. As the

TABLE 4-64. ESTIMATED CHARACTERISTICS OF FEED AND EFFLUENTS FOR THE COOLING TOWER OPERATION (1,134 Mg/hr Coal Feed EDS Plant)

Component	Base Case		MFS Case	
	Feed*	Drift/ Blowdown**	Feed*	Drift/ Blowdown**
H_2S	∿0	∿0	∿0	∿0
NH_3, mg/ℓ	10	40	10	40
Phenols, mg/ℓ	0.05-23	0.2-92	0.06-28	0.24-110
TOC, mg/ℓ	28-840	110-3,400	32-960	130-3,800
COD, mg/ℓ	44-3,600	170-14,000	50-4,000	200-16,000
Organic Acid, mg/ℓ	10-140	40-560	12-1,000	48-4,000
Flow rate, m^3/hr	260	0.65/65	210	0.53/53

*Feed ranges represented treated effluents from Function 4 and Function 5.

**Characteristics of Drift and Blowdown are assumed to be identical except the flow rates. For illustrative purposes here, the drift and blowdown composition and flow rates have been estimated assuming that all of the treated Stream A is fed to the cooling tower, that no cooling tower makeup is required beyond Stream A, and that the evaporation rate (as determined by temperature, humidity, etc.) is such that a 75% volume reduction is achieved through evaporation. In practice, a different plant water balance might be encountered, Stream A might be too small or too large to serve as total plant makeup, or evaporation rate might be different Under those conditions, the drift/blowdown composition would be different from that illustrated above.

slurry falls down through the tubes, part of the slurry water is vaporized by the steam condensing on the outside of the tubes. The vapor evaporated from the slurry is withdrawn through a mist eliminator to remove entrained droplets before entering a compressor suction line. The vapor steam enters the compressor where it is compressed about 14 to 56 kPa. This pressure raises the condensation temperature of the steam to about 3 to 12 K above the boiling temperature of the recirculating brine. The steam is then fed to the shell side of the evaporator tube bundle. As the steam condenses, its release of heat causes more water to evaporate on the inside of the tubes. The product water (condensed steam) is pumped from the evaporator shell through the feed reheater to return as much heat as possible to the process before it is recycled or sent to further treatment. The brine solids concentration in the sump is continuously monitored. To maintain a constant slurry concentration of about 200,000 to 400,000 mg/l total solids, a waste brine slurry is continuously removed.

Operation of this process is sensitive to dissolved gases and volatile organics in the feed since they only leave with the product water. When applied to Combined Steam A, Function 2 through 4 and maybe 5 controls would be required ahead of the evaporator to ensure the water recovered is of acceptable reusable quality. Table 4-65 presents the anticipated performance of the process. The feed composition range represents treated effluents from Function 4 and 5 controls. The TDS values are assumed to remain unchanged after the Function 2 through 5 controls.

The capital investment and total annualized costs for the two cases are summarized in Table 4-66. The detailed basis and assumptions for the estimates are presented in Appendix B-26. The capital investment ranges from 14 to 16 million dollars, or about 0.5% of the base plant cost. The total annualized cost ranges from 4.2 to 5 million dollars, or about \$2.4 to \$2.5/m^3 of water treated.

TABLE 4-65. ESTIMATED COMPOSITION OF FEED AND CONCENTRATE FOR THE VAPOR COMPRESSION EVAPORATION - STREAM A (1,134 Mg/hr Coal Feed EDS Plant)

Component	Base Case		MFS Case	
	Combined Stream Feed*	Concentrate** Effluent	Combined Stream Feed*	Concentrate** Effluent
H_2S ppm	~0	~0	~0	~0
NH_3 ppm	10	100	10	100
Phenol, ppm	0.05-23	0.5-230	0.06-28	0.6-280
TOC, ppm	28-840	280-8,400	32-960	320-9,600
COD, ppm	44-3,600	440-36,000	50-4,000	500-40,000
Organic Acid, ppm	10-140	100-1,400	12-1,000	120-10,000
TDS, ppm	4,000	40,000	4,900	49,000
Flow rate, m^3/hr	260	26	210	21

*Feed ranges represent treated effluent from Function 4 and 5.

**The concentrate flow rate is assumed to be 10% of the feed flow rate.

TABLE 4-66. SUMMARY OF COST ESTIMATES* FOR THE VAPOR COMPRESSION EVAPORATOR

Cost Item	Base Case	MFS Case
Capital Investment	$\$16 \times 10^6$	$\$14 \times 10^6$
Total Annualized Cost	$\$ 5.0 \times 10^6$	$\$ 4.2 \times 10^6$
% Base Plant Capital Cost	0.51%	0.49%
% Base Plant Annualized Cost	0.53%	0.47%
Unit Capital Investment	$62,000/$m^3$/hr	$67,000/$m^3$/hr
Unit Annualized Cost	$ 2.4/m^3	$ 2.5/m^3

* Capital investment is estimated using the procedures summarized in Table 4-1 and the installed equipment cost presented in Figure B 26-2. Annualized cost is estimated by using the bases presented in Table 4-2, and the utility and other requirements presented in Appendix B 26.

VCE generates: 1) a water product, 2) a small purge of noncondensible gases from the evaporator shell, and 3) a small waste brine (concentrate). The water product quality is dependent on the feed characteristics. Assuming the volatile components of the Combined Sour Water Stream are removed by Function 2 through 4 controls, the product water should essentially be void of any pollutants. Likewise, the purge from the evaporator shell should be negligible since all gases initially in the stream are stripped. The estimated characteristics of the concentrate is presented in Table 4-65.

4.3.1.1.6.3 Control Technique 3 - Reverse Osmosis (RO)

Reverse osmosis uses a semipermeable membrane and a pressure differential to separate relatively clean water from process wastewater. The pressure gradient forces the water through the membrane, leaving a concentrated solution of impurities.

The removal efficiency of specific ions by a reverse osmosis membrane can range from 90 to 95 percent of dissolved minerals, up to 99 percent of most dissolved organics greater than a molecular weight of 200, and up to 99% of biological and colloidal matter (94). Typically, about 75% of the feed is recovered as water. Since the detailed characteristics of Combined Stream A are not known, the actual performance of the process may differ from these estimates. Also, although the feed to the RO system is assumed to have gone through Function 4 or 5 treatments, these may still not be enough to prevent excessive plugging (due to complex organics and/or inorganic precipitates) of the membrane, making it nonfeasible to operate the process.

The capital investment for the base case (260 m^3/hr Stream A flowrate) and MFS case (210 m^3/hr) are estimated to be 2.3 and 1.9 million dollars, respectively; or about \$8,900/m^3/hr and \$9,100/m^3 of treated water (see Figure B22-1, Appendix B-22 for installed equipment cost and Table 4-1 for conversion to capital investment cost). The total annualized costs for the two cases are \$760,000/yr and \$640,000/yr, or about \$0.37/m^3 and \$0.39/m^3 of treated water (see Figure B22-2 and Table 4-2).

4.3.1.1.7 Control Function 7 - Final Disposal

Four final disposal alternatives for treated effluents are addressed in this manual. These alternatives are not intended to represent different levels of control, but rather fundamentally different approaches. Any of these approaches may be feasible alternatives for use at a given site depending on factors such as the availability of land, net evaporation rate, groundwater reservoir characteristics, and surface discharge restrictions.

4.3.1.1.7.1 Control Technique 1 - Surface Discharge

Discharge of treated effluent to surface water (stream, lake, ocean) is practiced by most industrial and municipal wastewater treatment plants. The Combined Sour Water Stream may require going through some or all of the functional controls to satisfy all applicable discharge requirements.

4.3.1.1.7.2 Control Technique 2 - Deep Well Injection

As potentially applied to Combined Stream A, deep well injection is essentially a "zero discharge" technique with the objective of maximizing water reuse. When deep well injecting any waste, the goal is to inject the waste into an underground formation without creating any adverse impacts on the already existing natural conditions. Selecting a zone with naturally high salt content would satisfy this goal. Other criteria for determining the feasibility of a disposal zone are summarized in Appendix B-27.

One major consideration in operating a deep well injection system is to pretreat the waste sufficiently well to avoid formation pluggage. It is assumed that Combined Stream A will undergo some form of volume reduction techniques in which water is recovered. The residual concentrated brine is then disposed to deep well. As shown in Tables 4-64 and 4-65, the organic loadings of the concentrated brine might be relatively high. Because of these high residual organic levels, the organics may need to be destroyed (e.g., in an incineration step following volume reduction) to prevent well formation pluggage.

Injection wells are generally between 300 and 1800 meters deep. Assuming an installed equipment cost for injection wells of \$328/m (see Appendix B-27 for basis and assumption), this would correspond to a capital investment of about \$180,000 to \$1 million (see Table 4-1 for bases for converting installed equipment cost to capital investment), or about \$8600/m^3/hr to \$48,000/m^3/hr for the MFS case and \$6,900/m^3/hr to \$38,000/m^3/hr for the base case.

4.3.1.1.7.3 Control Technique 3 - Surface Impoundment

In practice, the surface impoundment serves as a solar evaporation pond where all the water is evaporated. This is a viable final disposal alternative for southwestern U.S. plants where the arid conditions result in high net evaporation rates. To prevent excessive air emissions, the volative organics such as dissolved gases and organics such as phenols may need to be removed ahead of this technique. Depending on the amount of land available for use as impoundment, Combined Stream A may or may not need to go through any volume reduction step before being discharged to the impoundment. However, to maximize water reuse, it is assumed that Stream A will undergo some form of volume reduction step. The residual concentrated brine is then disposed to impoundment.

The costs of surface impoundments are largely costs associated with the purchase of land and the construction of the impoundment. The design and construction of the impoundment would aim at minimizing migration of leachate to groundwater and to surface waters. These would depend on site-specific factors such as local hydrogeologic conditions and net evaporation rates. Assuming a net evaporation rate of 0.5 m/yr (20 in/yr) and no volume reduction technique is applied, the capital investment for the base case (with pond area of 4.1×10^6 m^2) for a non-lined and a plastic-lined (30 mil Hypalon) impoundment with a 2.4 m dike is estimated to be \$15 x 10^6 and \$41 x 10^6, respectively;

for the MFS case (with pond area of 3.3×10^6 m^2), the equivalent costs are \$$6 \times 10^6$ and \$$33 \times 10^6$. The unit installed equipment cost for these estimates are presented in Figure B-28-5 (Appendix B-28) and the bases for converting installed equipment cost to capital investment are summarized in Table 4-1. The total annualized costs, which consisted of labor (assumed to be 2000 hr/yr), maintenance and amortized cost (see Table 4-2 for bases), are estimated to be \$$2.9 \times 10^6$/yr and \$$7.9 \times 10^6$/yr for the base case; and \$$1.8 \times 10^6$/yr and \$$64 \times 10^6$/yr for the MFS case.

4.3.1.1.7.4 Control Technique 4 - Co-Disposal with Ash

Co-disposal is one form of final disposal whereby the wastewater is used to moisturize the ash from the partial oxidation, the boiler or Flexicoking units. The wetted ash is then handled by a solid waste disposal method. As potentially applied to Stream A, co-disposal would be done only in the zero-discharge water reuse case. The feasibility of this technique is determined by the quantities of ash generated and its moisture holding capacity. Based on total feed coal rates of 1265 Mg/hr and 1296 Mg/hr for the base case and MFS case (see Table 3-3), and an ash content of 9.1%, about 120 Mg/hr of total ash will be generated. Assuming a moisture holding capacity of 20%, this would hold about 24 m^3/hr of water, i.e., Combined Stream A would need to be reduced to no greater than this amount, assuming this is the only stream from the plant to be disposed this way. As shown in Table 4-65, 90% evaporation of Stream A (e.g., through forced evaporation) can reduce the brine discharge rate to about this flow. However, as shown in Table 4-64, use of cooling tower provides less concentration; blowdown (brine) is greater than this flow.

It is assumed that co-disposal would not incur any incremental costs of solid waste disposal due to addition of brine. The solid waste disposal cost developed in Section 4.4 assumes that the ash would contain 20% moisture; thus, addition of brine up to this moisture level in the ash would not increase the cost of ash disposal in terms of the mass/volume of the material being disposed.

However, dry ash consists essentially of inorganic coal ash. If the brine is added, then the potential contaminants added by the brine must be considered in choosing and designing the solid waste disposal approach; this choice of disposal approach may increase the disposal costs.

4.3.1.2 Combined Stream B - Streams Containing Primarily Dissolved Gases

Combined Stream B consists of five streams for the base case and three streams for the MFS case. These are:

Base case

- Stream 312 - Flexicoking Heater Overhead Drum Wastewater
- Stream 403 - Knockout Drum Wastewater in H_2 Cryo Recovery
- Stream 430 - Blowdown and K.O. Drum Wastewater from H_2 Generation
- Stream 431 - Overhead Receiver Wastewater from Acid Gas Removal in H_2 Generation
- Stream 452 - Knockout Drum Wastewater in Ammonia Synthesis.

MFS case

- Stream 441 - Sour Water from Partial Oxidation Unit
- Stream 312 - Flexicoking Heater Overhead Drum Wastewater
- Stream 403 - Knockout Drum Wastewater in H_2 Cryo Recovery.

These streams are generally combined for treatment because they all contain dissolved gases, but little or no organics.

As discussed in Section 3, there is generally no quantitative information on the characteristics of the individual streams that make up Stream B. Thus, the dissolved gases concentrations in Stream B are not known. In general, these dissolved gases concentrations are expected to be fairly low. Nevertheless, it is assumed that one or more of these dissolved gases may exist in such a level in these streams that their removal from the streams may be desired. The controls applicable to Stream B are the Function 3 techniques (dissolved gas removal), which include, e.g., steam stripping. These techniques are discussed in Section 4.3.1.1.3.

In actual application, Stream B is likely to be combined with Stream A before the Function 3 technique in an integrated wastewater treatment system, in order to take advantage of the economy of scale. Also, combining the two streams may provide part or all of the dilution requirement for activated sludge treatments on Stream A in Function 4 (see 4.3.1.1.4.1). There are at least two possible ways of combining the two streams. In an intergrated control system where Function 2 (bulk organics removal e.g., solvent extraction) is placed ahead of Function 3 for Stream A, Stream B can be added to the system after Stream A has passed through Function 2. Alternatively in a scheme where Function 3 is ahead of 2, Stream B could undergo the stripping part of Function 3 independent of Stream A; bypass Function 2; and combine with Stream A ahead of Function 4 (see the integrated wastewater control examples in Section 4.3.1.6). Although it is assumed that these streams would be combined before the stripping unit, in reality separate strippers may be used for the streams which contain primarily ammonia (e.g., Stream 452) and streams which contain no ammonia. The ammonia stripper overhead would be routed for ammonia recovery while the overhead from the other stripper would be routed to acid gas removal system for further treatment.

Combining this stream with Stream A for treatment will incur additional incremental costs beyond those presented in Section 4.3.1.1 for Stream A alone. The total costs for treating these streams in integrated control systems are presented in Section 4.3.1.6.

4.3.1.3 Stream 451 - Aqueous Ammonia from Ammonia Synthesis

As discussed in 3.3.4.2.2, this stream is the product stream from ammonia synthesis operation in the base case design; it contains close to 12% ammonia. The ammonia can be recovered by any of the Function 3 techniques. As discussed in 4.3.1.1.3 and Appendix B-7, Function 3 techniques consist of a steam stripping unit and a recovery unit. It is likely that in an integrated wastewater treatment system, separate strippers will be used for Streams A, B and 451, but a common recovery unit will be designed to handle the combined stripper overheads. As discussed in 4.3.1.1.3, Function 3 techniques are expected to recover most of the ammonia in this stream, leaving a stripped effluent with about 50 to 200 mg/l ammonia. Combining this stream with Stream A for treatment will incur additional costs beyond the costs presented in Section 4.3.1.1 for Stream A alone. The total costs for treating these streams combined in integrated control systems are presented in Section 4.3.1.6.

Water Source Type 1
Slurry Dryer Cold
Separator Wastewater

4.3.1.4 Stream 103 - Slurry Dryer Cold Separator Wastewater

As presented in Table 3-10, this stream contains relatively low levels of dissolved gases (80 ppm NH_3, 4 ppm H_2S) and dissolved organics (1000 ppm phenol). This stream might be combined with Stream A ahead of Function 2 or Function 4 for treatment. Whether Function 2 (phenol recovery) is applied or not is a judgement which must be made on a case-by-case basis. One major factor affecting this judgement is the trade-off between the recovered phenol value and the incremental cost for the needed additional Function 2 treatment capacity to handle this stream. Combining this stream with Stream A for treatment will incur additional costs beyond the costs presented in Section 4.3.1.1 for Stream A alone. The total costs for treating these streams combined in integrated control systems are presented in Section 4.3.1.6.

4.3.1.5 Stream 443 - Slag Filtrate from Partial Oxidation Unit

This stream only exists in the MFS case. As presented in Table 3-52, this stream contains relatively low dissolved gases (about 130-380 ppm NH_3) and organics (about 28-355 ppm COD). This stream may be combined with Stream A ahead of Function 4 (e.g., biological oxidation) for treatment. The ammonia may serve as food to the microorganisms. Also, this may provide part or all of the dilution requirements for Function 4 control on Stream A (see 4.3.1.1.4.1). Combining this stream with Stream A for treatment will incur additional costs beyond the costs presented in Section 4.3.1.1 for Stream A alone. The total costs for treating these streams combined in integrated control systems are presented in Section 4.3.1.6.

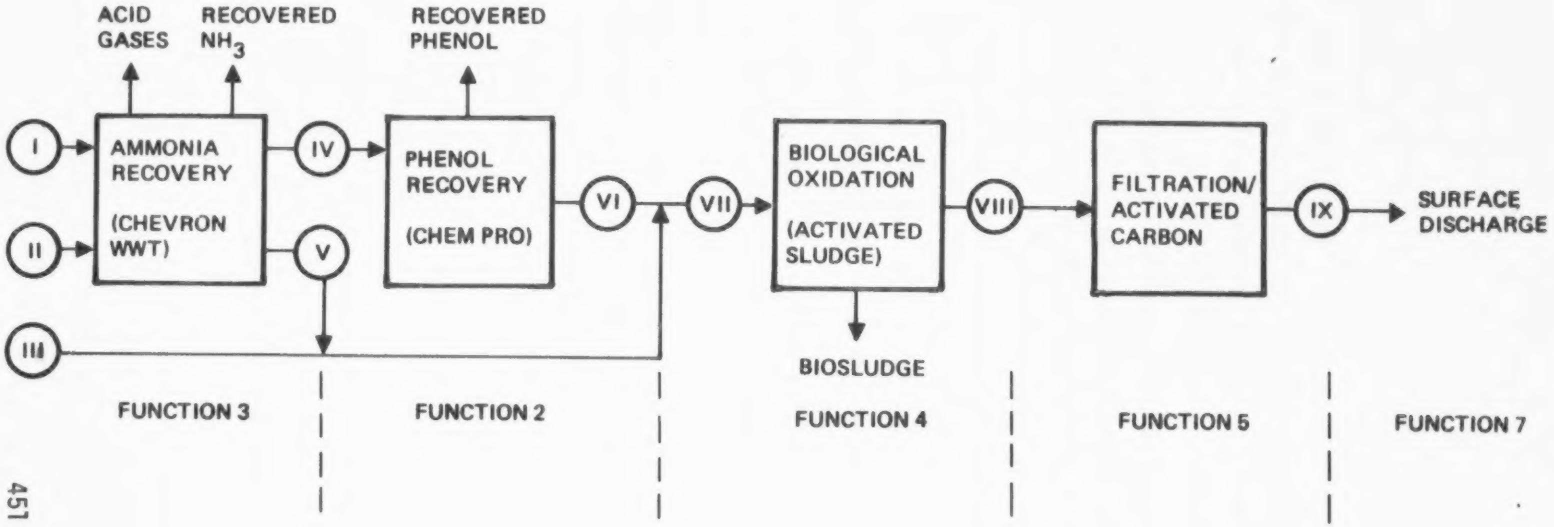
4.3.1.6 Integrated Control Examples for Treating Organics and Dissolved Gases Containing Wastewaters

As discussed previously in Section 4.3.1, there are 14 waste streams in the base case and 12 in the MFS case which contain high levels of organics or dissolved gases. Because of their similarity in characteristics and to take advantage of the economics of scale of a large, fully integrated treatment facility, these streams are expected to be combined at appropriate points for treatment.

This section presents examples of integrated controls for treating these streams. As discussed in previous sections, several treatment techniques would likely be combined for treating these streams, and these techniques can be combined in a large number of configurations to effect efficient control. The approaches considered in this section are by no means inclusive or indicative of necessary or sufficient treatment; they only serve as examples of how some available control techniques may be combined to control the subject waste streams.

4.3.1.6.1 Integrated Control Example 1: Chevron WWT + Chem-Pro + Activated Sludge + Activated Carbon + Discharge to Surface Water

Figure 4-4 presents a block flow diagram of an integrated control system for treating the Source Type 1 wastewaters. A total of five control functions are included in this example, namely, Removal of Bulk Organics (Function 2). Removal of Dissolved Gases (Function 3), Removal of Dissolved Organics (Function 4), Removal of Residual Organics (Function 5), and Final Disposal (Function 7). These functions are being considered by a number of proposed coal conversion facilities in the U.S., including EDS, SRC-II and Lurgi plants (93-98). The control techniques used in this example are Chem-Pro (Function 2), Chevron WWT (Function 3), Activated Sludge (Function 4), Activated Carbon (Function 5) and Surface Discharge (Function 7). The use of trade name processes such as Chem-Pro and Chevron WWT is simply to provide an example and is not intended as a preference or endorsement.


ACID GASES
RECOVERED NH_3
RECOVERED PHENOL
I
II
III
AMMONIA RECOVERY (CHEVRON WWT)
IV
V
PHENOL RECOVERY (CHEM PRO)
VI
VII
BIOLOGICAL OXIDATION (ACTIVATED SLUDGE)
BIOSLUDGE
VIII
FILTRATION/ ACTIVATED CARBON
IX
SURFACE DISCHARGE
FUNCTION 3
FUNCTION 2
FUNCTION 4
FUNCTION 5
FUNCTION 7

I = STREAM A (STREAMS 106, 152, 155, 202,
252, 307, 308)

II = STREAM B (STREAMS 312, 403, 430, 431, 452)
AND STREAM 451 FOR BASE CASE

= STREAM B (STREAMS 312, 403, 441)
AND STREAM 443 FOR MFS CASE

III = STREAM 103

Figure 4-4. Block Flow Diagrams for Integrated Control Example 1 for Treating Type 1 Wastewater

Water Source Type 1
Integrated Control
Example 1

As shown in Figure 4-4, the waste streams are combined into three groups with each group being treated by a different combination of techniques. Group I streams, which are the same as the Combined Stream A discussed in Section 4.3.1.1, includes the seven streams that contain high levels of NH_3, H_2S, phenols and other organics. This group enters the Chevron WWT System where H_2S, CO_2 and NH_3 are stripped. The H_2S/CO_2 stream is routed to acid gas recovery system for sulfur recovery, and the ammonia is recovered as anhydrous ammonia within the Chevron WWT System. The stripped wastewater passes on to the Chem-Pro System where phenol is recovered. The dephenolized wastewater goes through the activated sludge system where dissolved organics are removed. The activated sludge treated effluent is then passed through the activated carbon system for residual organic removal before discharge to surface water.

It is assumed that the relatively low concentrations of oil and grease in the stream will not affect the overall performance of the integrated control system.

Group II includes six streams for the base case (Stream B + 451) and four streams for the MFS case (Stream B + 443). These streams are believed to contain dissolved gases but low or no phenols. This group goes through a different stripper setup in the Chevron WWT System. The overheads are combined with those from treating the Group I streams for sulfur and ammonia recovery. The stripper bottom joins the dephenolized Group I streams for a biological and activated carbon treatment.

Group III consists of one stream which contains medium level of phenols (1000 ppm) which might not be economical to recover but is treatable by biological processes. This stream joins the other two groups ahead of the Activated sludge system.

The rationale for suggesting the ammonia stripping unit and phenol extraction for treating Group I streams, and the ammonia stripping unit for treating Group II streams ahead of the biological treatment unit is apparent. The phenol and ammonia levels in these waste streams might be toxic to the biota, or at least would render the bio-treatment unit extremely vulnerable to upset. In addition, recovered ammonia can be sold as a by-product, and recovered phenol can be used as fuel, if not for resale. It is estimated that 52,100 Mg/yr and 31,800 Mg/yr of ammonia, and 18,900 Mg/yr and 17,300 Mg/yr of phenol can be recovered from the base case and MFS case, respectively.

The optimal sequencing for the ammonia and phenol recovery processes cannot be assessed without detailed process information. In the commercial Lurgi gasification plant designs it has been proposed that the Lurgi raw gas liquor be first passed through an extraction process to recover phenol and then through a steam stripping unit to recover ammonia (95,96). The commercial SRC-II plant design (97) and the EDS plant design (98) proposed the sequencing as shown in Figure 4-4. It is anticipated that if phenol extraction were ahead of ammonia recovery, some of the dissolved gases would be stripped during the residual solvent recovery step of the extraction process (see Appendix B-5 for detailed process description). The stripped gas stream would be routed to the ammonia recovery unit. On the other hand, with the ammonia recovery ahead of phenol extraction, some phenols and other volatile organics will be stripped by the ammonia recovery strippers. These organics can be recovered by passing the stripper overhead through a condenser. The recovered organics stream is then fed to the solvent extraction unit. Regardless of the actual sequencing of the two units, the treated effluents from either system are expected to be similar. For the purpose of this document, the combined performances and the costs are assumed to be the same for either method of sequencing.

The estimated flow rates and compositions of the uncontrolled streams, treated effluent from each technique and the treated discharge stream for integrated example 1 are summarized in Tables 4-67 and 4-68 for the base case and MFS case, respectively. Detailed basis and assumptions for these estimates have been presented in the previous sections and the Appendices. The tables

TABLE 4-67. ESTIMATED COMPOSITION OF UNCONTROLLED WASTE STREAMS, TREATED EFFLUENTS AND SECONDARY WASTE STREAMS FOR INTEGRATED CONTROL EXAMPLE 1 BASE CASE (1,134 Mg/hr COAL FEED EDS PLANT)

Components, kg/hr	Stream I	Stream II	Stream III	Stream IV	Stream V	Stream VI	Stream VII	Stream VIII	Stream IX
H_2S	4,700	170	0.6	0	0	0	0.6	0	0
NH_3	3,600	3,000	12	13	7	13	22	5.5	5.5
CO_2	2,300	630	46	0	0	0	0	0	0
Phenols	2,400	0.5	150	2,400	0.5	24	170	9	0.9
Organic Acids	840	0	0	840	0	320	320	16	5
COD	12,000	2	720	12,000	2	4,800	5,500	1,100	110
TOC	3,700	0.8	230	3,700	0.8	1,100	1,300	260	26
Oil and Grease	50	0	0	50	0	0	0	0	0
TDS	1,000	90	0	1,000	90	1,000	1,100	1,100	1,100
Flow Rate	260,000	131,000	156,000	260,000	131,000	260,000	547,000	547,000	547,000

SECONDARY STREAMS

Component, kg.hr	Recovered Phenol	Recovered NH_3	Acid Gas	Biosludge
Flow Rate	2,400	3,600	7,000	4,200
NH_3	---	3,600	0.4	---
H_2S	---	0.2	4,700	---
H_2O	---	4	35	3,400
CO_2	---	0	2,300	---
Phenol	2,400	---	---	---
Organic Solids				800

TABLE 4-68. ESTIMATED COMPOSITION OF UNCONTROLLED WASTE STREAMS, TREATED EFFLUENTS AND SECONDARY WASTE STREAMS FOR INTEGRATED CONTROL EXAMPLE 1 MFS CASE (1,134 Mg/hr COAL FEED EDS PLANT)

Component Kg/hr	Stream I	Stream II	Stream III	Stream IV	Stream V	Stream VI	Stream VII	Stream VIII	Stream IX
H_2S	4,600	90	0.6	0	0	0	0.6	0	0
NH_3	3,500	540	12	11	4	11	27	4.4	4.4
CO_2	2,100	310	46	0	0	0	46	0	0
Phenols	2,200	0.2	150	2,200	0.2	22	170	9	0.9
Organic Acids	840	0	0	840	0	420	420	21	6
COD	11,000	50	720	11,000	50	4,400	5,200	1,000	100
TOC	3,400	2	230	3,400	2	1,000	1,200	240	24
Oil and Grease	50	0	0	50	0	0	0	0	0
TDS	1,000	90	nil	1,000	90	1,000	1,100	1,100	1,100
Flow Rate	210,000	71,000	156,000	210,000	210,000	210,000	437,000	437,000	437,000

SECONDARY STREAMS

Component, Kg/hr	Recovered Phenol	Recovered NH_3	Acid Gas	Biosludge
Flow Rate	2,200	3,500	6,700	3,300
NH_3	---	3,500	0.3	---
H_2S	---	0.2	4,600	---
H_2O	---	4	33	2,600
CO_2	---	0	2,100	---
Phenol	2,200	---	---	---
Organic Solids				700

also give the estimated flow rates for the secondary waste streams produced by this treatment scheme. Available controls for these secondary waste streams are discussed in the appropriate medium section. Not included in the tables is the flue gas generated during regeneration of the exhausted activated carbon. As discussed in 4.3.1.1.5.1 and Appendix B-15, it is not possible to determine the characteristics of this secondary stream. It is assumed that the organics adsorbed by the carbon will be destroyed during regeneration and the thermal regeneration unit is equipped with a scrubber to remove any particulate in the flue gas.

Tables 4-69 and 4-70 present, respectively, the estimated costs for integrated control example 1 for the base case and MFS case. The total capital investments are 50 and 43 million dollars, or about $\$98,000/m^3/hr$ and $\$92,000/m^3/hr$, expressed in terms of all treated streams combined (I, II and III). The total annualized costs are $10 million for both cases, or about $\$2.3/m^3$ and $\$2.9/m^3$ for the base case and MFS case, respectively. In all cases, the capital and annualized costs for each individual control technique in these tables are calculated in exactly the same manner as illustrated earlier for each technique as applied for Stream A alone (section 4.3.1.1). The costs in Tables 4-69 and 4-70 for each technique are sometimes higher than shown in the respective tables in Section 4.3.1.1 due to the increased flows resulting from the combination of other streams with Stream A as shown in Figure 4-4.

4.3.1.6.2 Integrated Control Example 2 - Chevron WWT + Chem-Pro + Activated Sludge + Activated Carbon + Cooling Tower + Surface Impoundment

Control Example 2, as shown in Figure 4-5, includes six techniques, namely, Chevron WWT, Chem-Pro, Activated Sludge, Activated Carbon, Cooling Tower and Surface Impoundment. The use of trade name processes such as Chem-Pro and Chevron WWT is simply to provide an example and is not intended as a preference or endorsement. Using treated process wastewater as make-up to cooling tower was one option considered by the ANG Lurgi gasification project at North

TABLE 4-69. SUMMARY OF ESTIMATED COSTS
FOR INTEGRATED CONTROL
EXAMPLE ONE-BASE CASE
(1,134 Mg/hr Coal Feed EDS Plant)

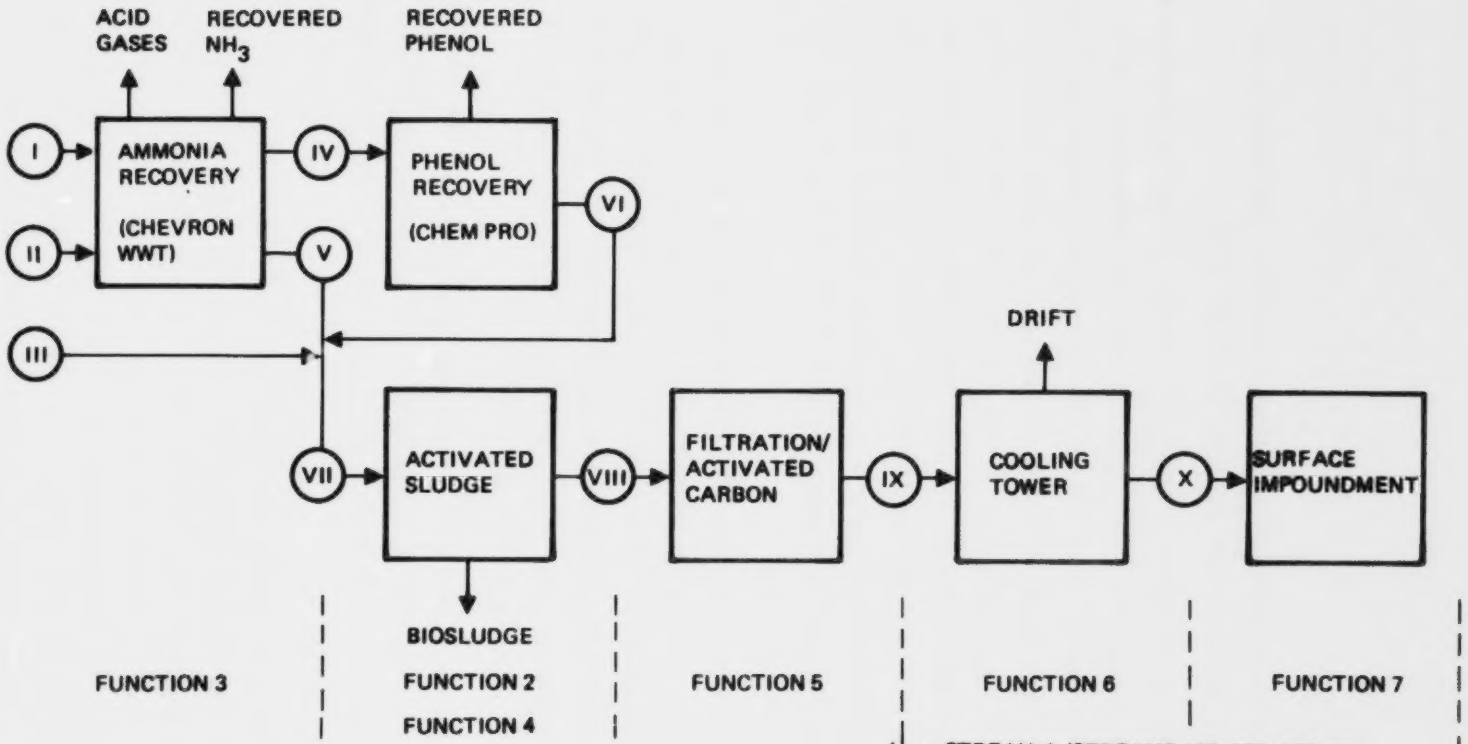
Item	Chevron WWT	Chem Pro	Activated Sludge	Activated Carbon	Total
Capital Investment, $\$10^6$	13	12.4	20	4.1	50
Total Annualized Cost, $\$10^6$/yr	0.73	2.6	4.9	1.9	10
Unit Capital Investment* $\$/m^3/hr$	24,000	23,000	37,000	7,500	92,000
Unit Annualized Cost*, $\$/m^3$	0.17	0.61	1.1	0.42	2.3
% Base Plant Investment,	0.41	0.39	0.63	0.13	1.6
% Base Plant Annualized Cost,	0.08	0.27	0.52	0.20	1.1

*Expressed in terms of m^3 of all treated waste streams combined. (I, II, and III).

TABLE 4-70. SUMMARY OF ESTIMATED COSTS FOR
INTEGRATED CONTROL EXAMPLE
ONE-MFS CASE
(1,134 Mg/Hr Coal Feed EDS Plant)

Item	Chevron WWT	Chem Pro	Activated Sludge	Activated Carbon	Total
Capital Investment, $\$10^6$	12	11	16	3.8	43
Total Annualized Cost, $\$10^6$/yr	1.8	2.2	4.2	1.8	10
Unit Capital Investment, $\$/m^3$/hr	27,000	25,000	37,000	8,700	98,000
Unit Annualized Cost, $\$/m^3$	0.52	0.64	1.2	0.52	2.9
% Base Plant Investment,	0.41	0.38	0.56	0.13	1.5
% Base Plant Annualized Cost,	0.20	0.25	0.47	0.20	1.1

*Expressed in terms of m^3 of all treated waste streams combined. (I, II and III).


ACID GASES
RECOVERED NH_3
RECOVERED PHENOL
I
II
III
AMMONIA RECOVERY (CHEVRON WWT)
IV
V
PHENOL RECOVERY (CHEM PRO)
VI
VII
ACTIVATED SLUDGE
BIOSLUDGE
VIII
FILTRATION/ ACTIVATED CARBON
IX
DRIFT
COOLING TOWER
X
SURFACE IMPOUNDMENT
FUNCTION 3
FUNCTION 2
FUNCTION 4
FUNCTION 5
FUNCTION 6
FUNCTION 7

I = STREAM A (STREAMS 106, 152, 155, 202, 252, 307, 308)

II = STREAM B (STREAMS 312, 403, 430, 431, 452) AND STREAM 452 FOR BASE CASE

= STREAM B (STREAMS 312, 403, 441) AND STREAM 443 FOR MFS CASE

III = STREAM 103

Figure 4-5. Block Flow Diagram for Integrated Control Example 2 for Treating Type 1 Wastewater

Dakota (91); disposing treated process wastewater in surface impoundment was considered by El Paso Lurgi gasification project at Texas (96).

This example is likely to be feasible only for plants in the southwestern part of the U.S. where net evaporation is high; the surface impoundment will serve as a solar evaporation pond. Passing the treated water through cooling tower before discharging to the pond will reduce raw cooling water requirement which is likely an important consideration in the southwestern states. This treatment also reduces the area required for the impoundment, since only the concentrated blowdown from the cooling tower will go to the impoundment. The four techniques preceding the cooling tower will minimize emission of volatile gases and organics through the cooling tower, and minimize cooling tower failure due to pluggage and other problems.

The characteristics of the waste streams I through IX, as well as the secondary streams Acid Gases, Recovered Phenol, Recovered NH_3 and Biosludge, are identical to those presented in Tables 4-67 and 4-68. The characteristics of stream X, the concentrated blowdown from the cooling tower is presented in Table 4-71. It is assumed the cooling tower is operating at a concentration cycle of four, and no other make-up water is required.

This example generates one additional secondary stream beyond the previous example. This is the drift from the cooling tower. The characteristics of this stream are presented in Table 4-71. The basis for this estimate have been presented in Section 4.3.1.1.6.1.

The estimated cost for this system is summarized in Table 4-72. The first four treatment techniques are similar to control example 1, and the costs of these have been presented before. The cooling tower is considered as part of the base plant operation, hence, no cost estimate is presented. The surface impoundment capital investments are estimated to be 7.8 and 39 million dollars for the base case; 6.2 and 31 million dollars for the MFS case, depending on whether the impoundment is lined or not (see Appendix B-24 for basis and

TABLE 4-71. COMPOSITION OF TREATED EFFLUENT FOR INTEGRATED CONTROL EXAMPLE 2 BASE CASE AND MFS CASE (1,134 Mg/hr Coal Feed EDS Plant)

Component, Kg/hr	Base Case*			MFS Case**		
	Stream IX	Stream X	Drift	Stream IX	Stream X	Drift
H_2S	0	0	0	0	0	0
NH_3	5.5	5.5	0.01	4.4	4.4	0.01
CO_2	0	0	0	0	0	0
Phenols	0.9	0.89	0.009	0.9	0.89	0.009
Organic Acids	5	4.95	0.05	6	5.94	0.06
COD	110	109	1.1	100	99	1
TOC	26	25.7	0.26	24	23.8	0.24
Oil and Grease	0	0	0	0	0	0
TDS	1,100	1,090	10	1,100	1,090	10
Flow Rate	547,000	137,000	1,370	437,000	109,000	1,090

*Characteristics of Streams I through VIII and other secondary streams presented in Table 4-67.

**Characteristics of Streams I through VIII and other secondary streams presented in Table 4-68.

TABLE 4-72. SUMMARY OF COSTS FOR INTEGRATED CONTROL EXAMPLE 2 (1,134 Mg/hr Coal Feed EDS Plant)

Item	Base Case			MFS Case		
	Other* Processes	Surface** Impoundment	Total	Other* Processes	Surface** Impoundment	Total
Capital Investment, $\$10^6$	50	7.8-39	58-89	43	6.2-31	49-74
Total Annualized Cost, $\$10^6$/yr	10	1.6-7.8	12-18	10	1.3-6.2	12-16
Unit Capital Investment,*** $\$/M^3$/hr	92,000	14,000-71,000	106,000-161,000	98,000	14,000-71,000	112,000-169,000
Unit Annualized Cost,*** $\$/m^3$	2.3	0.37-1.8	2.7-4.1	2.9	0.38-1.8	3.3-4.7
% Base Plant Capital Investment, %	1.6	0.24-1.2	1.8-2.8	1.5	0.22-1.1	1.7-2.6
% Base Plant Cost, %	1.1	0.17-0.82	1.3-1.9	1.1	0.15-0.70	1.3-1.8

*Costs for the individual processes preceding surface impoundment are presented in Tables 4-69 and 4-70.

**Lower range values are for no-liner impoundment and higher range values are for plastic-lined impoundment.

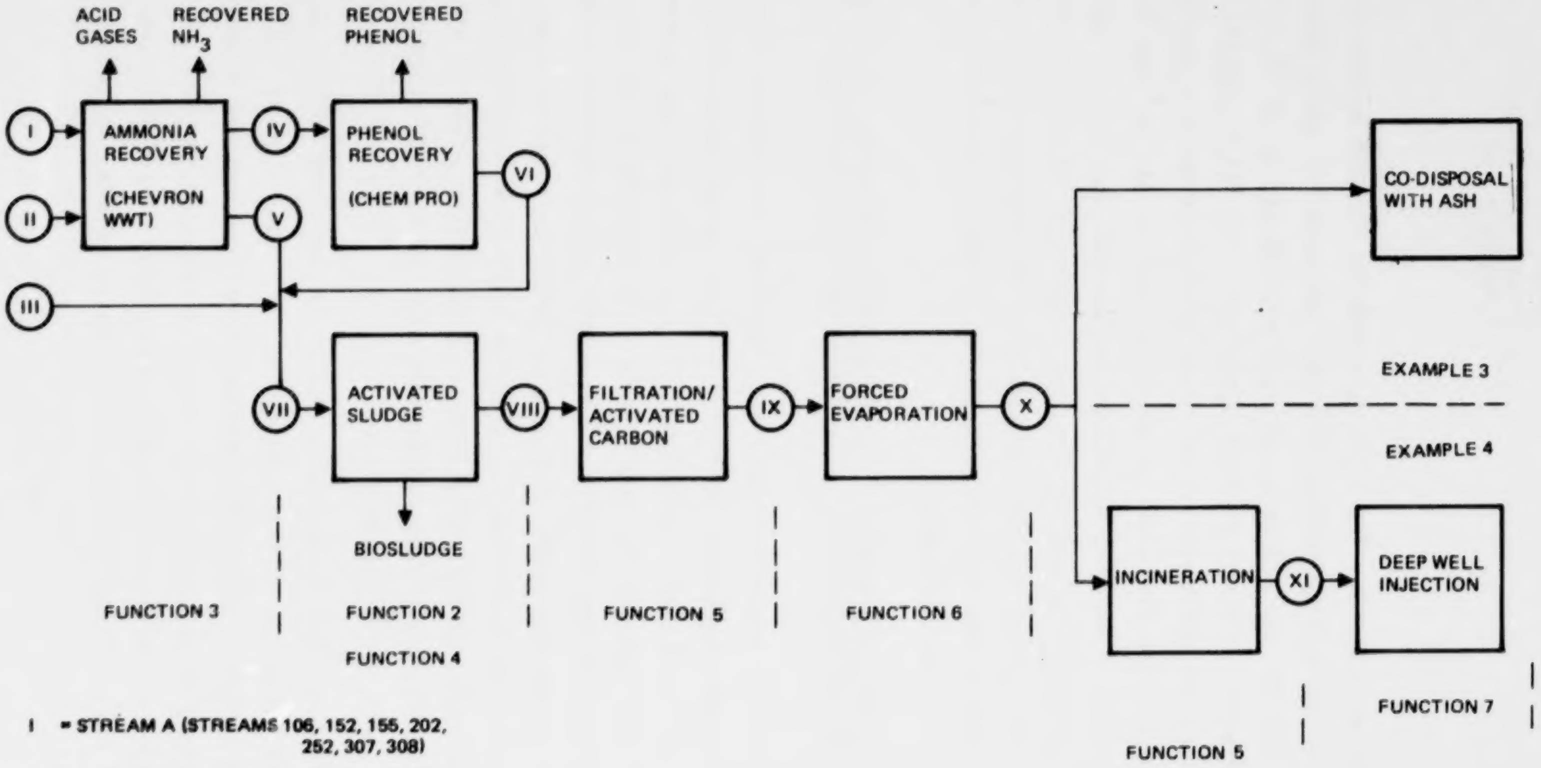
***Expressed in terms of m^3 of all treated waste streams combined (I, II and III).

assumptions for cost estimates). The total investments for the system are estimated to be 58 to 89 million dollars for the base case, or about \$106,000 to \$161,000/m^3/hr; and for the MFS case, these are estimated to be 49 to 74 million dollars, or about \$112,000 to \$169,000/m^3/hr. The total annualized cost for the two cases are estimated to be 12 to 18 million dollars per year and 12 to 16 million dollars per year, or about \$2.7/m^3 to \$4.1/m^3 and \$3.3/m^3 to \$4.7/m^3, respectively. The cost estimates for surface impoundment shown in Table 4-72 were estimated in exactly the same manner as illustrated in Section 4.3.1.1.7.3 for Stream A alone, except adjusted downward to account for the lower flow rate of the combined stream due to the volume reduction technique used ahead of the impoundment.

4.3.1.6.3 Integrated Control Example 3 - Chevron WWT + Chem-Pro + Activated Sludge + Activated Carbon + Forced Evaporation + Co-Disposal with Ash

Control Example 3, as shown in Figure 4-6, includes six control techniques, namely, Chevron WWT, Chem-Pro, Activated Sludge, Carbon Adsorption, Forced Evaporation and Co-Disposal with Ash. The use of trade name processes such as Chevron WWT and Chem-Pro is simply to provide an example and is not intended as a preference or endorsement. This scheme would optimize water reuse in the plant by recovering water through the forced evaporation step. The concentrated brine from the evaporator is used to moisturize (quench) ashes from the Flexicoking unit, partial oxidation unit and boiler prior to ash disposal. Disposal of concentrate from evaporator with ash was one scheme considered by the ANG Lurgi gasification project.

As discussed in 4.3.1.1.7.4, assuming an average moisture holding capacity of 20% for the ashes, about 24 m^3/hr of water can be disposed of as part of the ash solid waste stream. Forced evaporation (vapor compression evaporator) is used in this example to concentrate the waste stream down to this level. (A more economic alternative may be to use a combination of cooling tower, reverse osmosis and forced evaporation). As discussed before, to ensure proper


ACID GASES
RECOVERED NH_3
RECOVERED PHENOL
I
II
III
AMMONIA RECOVERY (CHEVRON WWT)
IV
V
PHENOL RECOVERY (CHEM PRO)
VI
VII
ACTIVATED SLUDGE
BIOSLUDGE
VIII
FILTRATION/ ACTIVATED CARBON
IX
FORCED EVAPORATION
X
CO-DISPOSAL WITH ASH
EXAMPLE 3
EXAMPLE 4
INCINERATION
XI
DEEP WELL INJECTION
FUNCTION 3
FUNCTION 2
FUNCTION 4
FUNCTION 5
FUNCTION 6
FUNCTION 5
FUNCTION 7

I = STREAM A (STREAMS 106, 152, 155, 202, 252, 307, 308)

II = STREAM B (STREAMS 312, 403, 430, 431, 452) AND STREAM 452 FOR BASE CASE

= STREAM B (STREAMS 312, 403, 441) AND STREAM 443 FOR MFS CASE

III = STREAM 103

Figure 4-6. Block Flow Diagram for Integrated Control Examples 3 and 4 for Treating Type 1 Wastewater

operation of the evaporator and that the water recovered is of reasonable qualities, volatile species and other pollutants would be removed ahead of the evaporator. This is accomplished by the four techniques shown in Figure 4-6 preceding forced evaporation. The estimated characteristics of the effluent from the evaporator are presented in Table 4-73. As shown, the streams will need to be concentrated about 24 and 18 times for the base case and MFS case, respectively. Only streams IX and X, which are directly related to the evaporator, are included in the table. The characteristics of other streams, including the secondary streams, have been presented in Tables 4-67 and 4-68. No new secondary waste stream is expected from the evaporator and the co-disposal operations.

The estimated costs for the evaporator and the total system are summarized in Table 4-74. The costs for the system include costs for Chevron WWT, Chem-Pro, Activated Sludge and Activated Carbon which have been presented in the previous examples, and forced evaporation. The co-disposal cost is assumed to be part of the solid waste disposal cost, and is not included here (see Section 4.3.1.1.7.4). Costs for forced evaporator of the combined stream in Table 4-74 are estimated in exactly the same manner as in Section 4.3.1.1.6.2 for stream A alone, except adjusted for the larger flow rate of the combined stream.

The concentrate from the evaporator contains fairly high concentration of organics which may be candidate for removal before final disposal with the ash. One applicable control technique is incineration. Incineration, if added to the two cases, will add 4.8 million dollars to the total capital investment and 4.8 million per year to the total annualized cost (see Table 4-76 for details). Removing the organics before disposal may reduce the solid waste disposal facility design/operation requirements and hence the total disposal cost, i.e., there may be a tradeoff in costs between the water and solid waste media for this control scheme.

TABLE 4-73. ESTIMATED COMPOSITION OF TREATED EFFLUENTS FOR INTEGRATED CONTROL EXAMPLE 4 BASE CASE AND MFS CASE (1,134 Mg/hr Coal Feed EDS Plant)

	Base Case*		MFS Case**	
Components, Kg/hr	Stream IX	Stream X	Stream IX	Stream X
H_2S	0	0	0	0
NH_3	5.5	5.5	4.4	4.4
CO_2	0	0	0	0
Phenols	0.9	0.9	0.9	0.9
Organic Acids	5	5	6	6
COD	110	110	100	100
TOC	26	26	24	24
Oil and Grease	0	0	0	0
TDS	1,100	1,100	1,100	1,100
Flow Rate	547,000	24,000	437,000	24,000

*Characteristics of Streams I through VIII and secondary streams are presented in Table 4-67.

**Characteristics of Stream I through VIII and secondary streams are presented in Table 4-68.

TABLE 4-74. SUMMARY OF ESTIMATED COSTS FOR
INTEGRATED CONTROL EXAMPLE 3
BASE CASE AND MFS CASE
(1,134 Mg/hr Coal Feed EDS Plant)

Item	Base Case			MFS Case		
	Other Processes*	Evaporator	Total	Other Processes*	Evaporator	Total
Capital Investment, $\$10^6$	50	26	76	43	22	65
Total Annualized Cost, $\$10^6$/yr	10	8.6	19	10	7.2	17
Unit Capital Investment,** $\$/m^3/hr$	92,000	48,000	140,000	98,000	50,000	148,000
Unit Annualized Cost,** $\$/m^3$	2.3	2.0	4.3	2.9	2.1	5.0
% Base Plant Capital Investment, %	1.6	0.82	2.4	1.5	0.76	2.3
% Base Plant Annualized Cost %	1.1	0.91	2.0	1.1	0.81	1.9

*Costs for individual processes are in Tables 4-69 and 4-70.

**Expressed in terms of m^3 of all treated waste streams combined (I, II and III).

4.3.1.6.4 Integrated Control Example 4 - Chevron WWT + Chem-Pro + Activated Sludge + Activated Carbon + Forced Evaporation + Incineration + Deep Well Injection

Control Example 4, as shown in Figure 4-6, includes seven control techniques, namely, Chevron WWT, Chem-Pro, Activated Sludge, Carbon Adsorption, Forced Evaporation, Incineration and Deep Well Injection. The use of trade name processes such as Chevron WWT and Chem-Pro is simply to provide an example and is not intended as a preference or endorsement. Deep Well injecting the concentrate from evaporation was one scheme considered by the ANG Lurgi gasification project (91). Incineration was included to reduce the threat of well plugging by organics in the brine. This example would optimize water reuse by recovering water through the evaporation step, and would be applicable to sites where local hydrogeological conditions permit deep well injection of the treated wastewaters.

The estimated characteristics of the treated streams for this system are presented in Table 4-75. Only streams X and XI, which are directly related to the incineration operation are included in the table. Characteristics of streams I through VIII and the secondary waste streams have been presented in Tables 4-67, 4-68 and 4-73. As shown, incineration is expected to remove 99.9% of the organics.

In addition to the recovered NH_3, recovered phenol, acid gases and biosludge, this system generates another secondary waste stream, namely, the offgas from the incinerator. As discussed in 4.3.1.1.5.3 this stream may contain SO_2, NO_x and particulates. The major contribution to the SO_2 and NO_x is likely from the fuel used. The particulates may come from the fuel and the wastewater. Since control for this stream is considered as part of the incinerator system*

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* Of the 83 installed incinerators designed to burn aqueous waste containing sodium salts, 80 have a quench tank followed by a venturi scrubber, 3 have electrostatic precipitators (91).

TABLE 4-75. ESTIMATED COMPOSITION OF TREATED EFFLUENTS FOR INTEGRATED CONTROL EXAMPLE 4 BASE CASE AND MFS CASE*
(1,134 Mg/hr Coal Feed EDS Plant)

Component, Kg/hr	Stream X	Stream XI	Stream X	Stream XI
H_2S	0	0	0	0
NH_3	5.5	5.5	4.4	4.4
CO_2	0	0	0	0
Phenols	0.9	0.0009	0.9	0.0009
Organic Acids	5	0.005	6	0.006
COD	110	0.11	100	0.1
TOC	26	0.026	24	0.024
Oil and Grease	0	0	0	0
TDS	1,100	1,100	1,100	1,100
Flow Rate	24,000	24,000	24,000	24,000

*Characteristics of Streams I through IX are presented in Tables 4-67 and 4-68.

no separate estimates on the characteristics or cost are presented for this stream.

The estimated costs for this control example are summarized in Table 4-76. As shown, the total capital investments for the base case and MFS case are estimated to be 82 and 71 million dollars, respectively; and the annualized costs are 25 and 23 million dollars per year, respectively. These correspond to \$150,000/m^3/hr and \$161,000/m^3/hr, and \$5.7/m^3 and \$6.7/m^3. Costs for incineration and deep well injection of the combined stream in Table 4-76 are estimated in exactly the same manner as in Section 4.3.1.1.5.3 and 4.3.1.1.7.2 for Stream A alone, except adjusted for the larger flow rates of the Combined stream.

TABLE 4-76. SUMMARY OF ESTIMATED COSTS FOR INTEGRATED CONTROL EXAMPLE 4 BASE CASE AND MFS CASE (1,134 Mg/hr COAL FEED EDS PLANT)

	Base Case				MFS Case			
	Other Processes*	Incineration	Injection	Total	Processes	Incineration	Injection	Total
Capital Investment, $\$10^6$	76	4.8	0.77	82	65	4.8	0.77	71
Total Annualized Cost, $\$10^6$/yr	19	4.8	1.1	25	17	4.8	1.1	23
Unit Capital Investment,** $\$/m^3/hr$	140,000	9,000	1,000	150,000	148,000	11,000	2,000	161,000
Unit Annualized Cost,** $\$/m^3$	4.3	1.1	0.25	5.7	5.0	1.4	0.32	6.7
% Base Plant Capital Investment,	2.4	0.15	0.2	2.6	2.3	0.17	0.03	2.5
% Base Plant Annulaized Cost,	2.0	0.51	0.12	2.6	1.9	0.54	0.12	2.6

*Cost for the individual processes preceding incineration are presented in Tables 4-69, 4-70, and 4-74.

**Expressed in terms of all treated waste streams combined (I II and III).

4.3.2 Source Type 2 - Inorganic Containing Wastewaters

Streams that fall into this source type include:

Stream No.	Stream Description
012	Coal Pile Runoff
702	Blowdown from Steam Generation System
708	Blowdown from Power Generation System
723	Regeneration Wastes from Water Demineralization
732	Cooling Tower Blowdown

These streams are common to both the base case and the MFS case, but are not unique to the EDS process. They are common to other industrial facilities such as oil- and coal-fired power generation plants and refineries. Existing and proposed regulations on the Electric Power Industry specify discharge limits on total suspended solids, total residual chlorine, and pH from these streams. These regulations also specify no discharge of priority pollutants resulting from chemical additives.

Table 4-77 summarizes the control functions and techniques that are applicable to these streams. As shown, control functions include Removal of Suspended Solids, Tars and Oils, Removal of Dissolved Inorganics, Volume Reduction and Final Disposal. Although listed as a separation function, Removal of Dissolved Inorganics is typically used in conjunction with Removal of Suspended Solids, Tars and Oils.

TABLE 4-77 SUMMARY OF CONTROL TECHNIQUES POTENTIALLY APPLICABLE TO EDS WASTEWATER CONTAINING PRIMARILY DISSOLVED INORGANICS

Table 4-45. SUMMARY OF CONTROL TECHNIQUES POTENTIALLY APPLICABLE TO THE TREATMENT OF EDS WASTEWATER CONTAINING PRIMARY DISSOLVED ORGANICS AND DISSOLVED GASES

Technology	Technology Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-Products and Secondary Waste Streams	Comments
Removal of Suspended Solids, Tars and Oils						
Gravity separation - API oil separator - Parallel plate	Provision of adequate residence time in a stagnant vessel to allow suspended solids or immiscible fluids to separate into lighter and heavier than water phases.	Suspended solids, tars and oils.	Depends upon design. 10-50% removal of TSS typical, 60-99% for oils.	Minimum feed stream turbulence.	Recovered oils (Sp. gr. < 1), sludges/solids (Sp. gr. >1)	Incorporated into the tar/oil separation system design and the biological treatment system.
Coagulation/ flocculation	Use of agents to promote the coalescence of fine suspended solids and adsorption of tars and oils; generally used in conjunction with a gravity separation process.	Promotes removal of finely dispersed particles.	Outlet suspended solids concentration to 10 mg/ℓ possible, oils removal of 60-95%.	A wide range of commercial flocculants are available.	Same as gravity separation.	Widely used in water treatment system to remove fine soilds.
Air flotation - Dissolved air - Induced air	Use of air bubbles to promote the disengagement of lighter-than-water materials from solution.	Suspended oils and solids.	Depends on characteristics of source and treatment process; TSS removal of 20-75%, oil removals of 75-85%.	Air requirements depend upon waste characteristics.	Recovered oils entrained, solids.	Widely used in various industries.
Filtration	Passing wastewater through suitable filter medium; filter material discarded or cleaned by backflushing.	Depends on filter medium; both coarse and fine structure materials are used industrially.	TSS removals of 30-90+%, oil removals of 65-90%	Filter media (sand, clay, fabric or polymeric membrane).	Filter backwash; spent filter media.	Proposed for use as polishing step for phenolic water downstream of tar/oil separation. "Sticky" tars/oils may cause problems with filter plugging and regeneration.

(Continued)

TABLE 4-77. (CONTINUED)

Technology	Technology Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-Products and Secondary Waste Streams	Comments
Removal of Dissolved Inorganics						
Ion exchange	Substitution of H^+, Na^+, OH^- or Cl^- ions for other ionic species, exchange resins regenerated with acid, base or salt solutions	Heavy metal, F^-, CN^-, scaling species NH_3	90⁺% removal for most ions, regeneration frequency is a key parameter	Regenerants, replacement resins	Spent regenerants and resins, treated water	Most effective as a polishing process. Clearly applicable to boiler feedwater treatment needs. Of limited use in treating process wastewaters containing high concentrations of organics or dissolved solids.
Chemical precipitation	Use of agents to promote the precipitation of inorganic solids from wastewaters	Ca, Mg, heavy metals, alkalinity	Varies with wastestream constituents. Typical removals: Cd 2% Ni 50% Cr 40% Pb 5% Cu 20% Se 10% Hg 20% Zn 25%	Lime, polymer and soda ash may be required.	Sludge contaminated with heavy metals	Generally followed by filtration and/or activated carbon adsorption.

TABLE 4-77. (CONTINUED)

Technology	Technology Principle	Components Removed	Removal Efficiency	Feed Requirements/ Restrictions	By-Products and Secondary Waste Streams	Comments
Volume Reduction						
Membrane separation - reverse osmosis	Use of semipermeable membrane and pressure to separate water from its dissolved constituents	Relative rejection efficiencies of the various soluble species will be determined by membrane characteristics and conditions of process operation	90-95% rejection of dissolved salts. Reductions in dissolved organics and BOD of up to 99%	Membrane	Spent membrane material, recovered water, brine	May be useful as preconcentration step prior to further treatment or ultimate disposal of wastewater. Membrane scaling and fouling with organics may limit the applicability of this technology to the treatment of process condensates.
- electrodialysis	Use of selective anion- or cation-permeable membranes with electric field to separate mineral ions from water					
Forced evaporation	Thermally induced evaporation of excess wastewater, condensate recovery optional	All nonvolatile species will remain in brine		None	Recovered condensate, noncondensible gases, waste brine	Very stringent materials requirements.
Cooling tower concentration	Wastewater used as partial makeup to the cooling tower and thereby concentrated into the blowdown	All nonvolatile species concentrated into the blowdown		Feed characteristics limited by corrosion, scaling, and biological fouling	Evaporation/ drift	
Residual Disposal						
Surface discharge	Wastes are conveyed to and mixed with a natural water source	Entire stream	100%	Restrictions are site-specific	None	
Deep well injection	Wastes are pumped into subsurface geological formations where they are isolated from all surface and groundwater supplies	Entire stream	100%	Injected fluids must be filtered to 5 micrometers and have a low organic content to prevent plugging. Wastes must not precipitate in the well or when mixed with subsurface fluids. Volume reduction prior to injection is often economical	None	
Surface impoundment	Wastes are held in a containment basin	All nonvolatile species remain as residual in the impoundment	100%	Concentrations of volatile species may need to be low to prevent loss to the atmosphere	Loss of volatile species	This technique is limited to locations having available land and net evaporation rates exceeding 0.5 m/yr (20 in/yr).
Co-disposal	Ash is quenched with wastewater then handled by a solid waste disposal technique	Entire stream	100%	Concentrations of volatile species may need to be low to prevent loss to the atmosphere	Loss of volatile species	

Water Source Type 2
Combined Inorganic Stream
Control Function 1

4.3.2.1 Combined Inorganic Stream - Streams 702, 708, 723 and 732

Combined inorganic stream consists of four streams, namely,

- Stream 702 - Blowdown from Steam Generation System
- Stream 708 - Blowdown from Power Generation System
- Stream 723 - Regeneration Wastes from Water Demineralization
- Stream 732 - Cooling Tower Blowdown.

As discussed in Section 3, these streams are not unique to the EDS plant. They are common to all steam electric power plants. It is expected that these streams will contain suspended solids and may be acidic. To be consistent with the discharge standards for steam electric power plants, it is assumed that biocides and anticorrosion agents used in cooling tower do not contain any priority pollutants such as chromium, zinc and nickel.

4.3.2.1.1 Control Function 1 - Removal of Suspended Solids, Tars and Oils

Two techniques may be used to treat these streams. These are presented below.

4.3.2.1.1.1 Control Technique 1 - Gravity Separation

This technique operates under the principle that by providing enough residence time in a quiescent reactor, the suspended solids will be separated from the wastewater due to their difference in density. The reactor may be a pond (surface impoundment), or a rectangular or circular tank. In the existing steam electric power plants, this technique typically produces an effluent with less than 30 ppm suspended solids.

The combined stream flow rates for the base case and MFS case were estimated to be, respectively, 730 m^3/hr and 600 m^3/hr (see Tables 3-76 and 3-77 for flow rates of the individual streams). Assuming a pond depth of 10 m and residence time of 24 hrs (to provide some capacity for holding storm runoffs), treating this stream in a surface impoundment would require impoundment surface area of 1750 m^2 and 1440 m^2 for the base case and MFS case, respectively. From

Figure B-28-1 in Appendix B-28 (for installed equipment cost) and Table 4-1 (for converting installed equipment cost to capital investment) and assuming the impoundment has a 2.4 m dike, the capital investments for the base case are estimated to be $35,000 and $150,000, respectively, for an unlined and Hypalon-lined impoundment; for the MFS case, these are, respectively, $31,000 and $130,000. The total annualized costs which are consisted mainly of labor (assumed to be 2000 hr/yr), maintenance and amortized costs (see Table 4-2 for bases) are estimated to be $43,000 and $65,000 for the base case and $42,000 and $61,000 for the MFS case. All these costs are less than 0.01% of the uncontrolled base plant cost.

Secondary Stream

This technique generates a secondary waste stream, namely, the settled sludge. There are no data to characterize this stream. In a surface impoundment, the settled sludge may be dredged from time to time, and disposed in a landfill, or be left in the impoundment permanently. The costs estimated in the previous paragraph do not include any closure cost or dredging cost.

4.3.2.1.1.2 Control Technique 2 - Filtration

In filtration, wastewater enters the filtration unit and slowly percolates through the filter media to the underdrain. Suspended materials are trapped in the filter media and are later flushed out of the filter during the backwash cycle. This technique typically reduces suspended solids to below 10 ppm.

The capital investments for this process are estimated to be $\$1.3 \times 10^6$ and $\$1.1 \times 10^6$, respectively, for the base case and MFS case (see Figure 8-4-1 in Appendix B-4 for unit installed equipment cost and Table 4-1 for bases for converting installed equipment cost to capital investment); or less than 0.1% of the uncontrolled base plant cost. The total annualized costs are estimated to be $380,000/yr and $320,000/yr for the base case and MFS case, respectively; or less than 0.1% of the uncontrolled base plant annualized cost.

Secondary Stream

Filtration generates one secondary waste stream, namely, the filter backwash. This stream is generally routed to sedimentation unit for treatment.

4.3.2.1.2 Control Function 2 - Removal of Dissolved Inorganics

Dissolved inorganics removal techniques are designed to neutralize the waste streams, to remove toxic trace elements or to lower the calcium, magnesium, silica, carbonate and/or sulfate levels. The latter species form the majority of the compounds which cause plugging, scaling and fouling problems when their solubility products are exceeded.

4.3.2.1.2.1 Control Technique 1 - Neutralization

This technique involves addition of chemicals (acids or bases) to neutralize the stream. The chemicals are typically either added prior to Function 1 technique or just before discharge. Equipment required for this technique includes chemical storage and feed system. Unless the waste streams are extremely acidic or basic, costs for these equipments and the chemical costs are expected to be insignificant when compared to other control technique costs. No secondary waste stream is generated from this process.

4.3.2.1.2.2 Control Technique 2 - Chemical Precipitation

Chemical precipitation involves the addition of chemicals, usually lime and possibly flocculant aide whereby the wastewater pH is raised and metal oxide/hydroxide precipitates are formed. The precipitates are removed by gravity settling. Depending upon the resultant pH, the treated waste may be recarbonated to lower the pH and to remove excess calcium (as $CaCO_3$) in a second stage settling tank. The process is usually followed by filtration to remove residual solids. When applied to the Combined Inorganic Stream, this technique is expected to be used in conjunction with one of the Function 1 techniques (see Section 4.3.2.1.1). Equipment required in addition to those already existing in Function 1 technique include chemical feed and chemical storage system. The costs for these equipments and the chemical costs for treating these streams are expected to be minor when compared to the Function 1 techniques costs.

Secondary Stream

This process generates a secondary stream, namely, the settled precipitates. This stream may be disposed with other inorganic solid waste streams generated from the plant.

4.3.2.1.3 Control Function 3 - Volume Reduction

Techniques considered in this function are similar to those discussed under Section 4.3.1.1.6 for Water Source Type 1 controls. The performances are expected to be similar whether the techniques are applied to Water Source Type 1 or Type 2 waste streams, although pretreatment requirements are different (e.g., no dissolved gases or bulk organic removal techniques are required for this stream). The costs and performances for these techniques have been presented in Section 4.3.1.1.6 and the appendices.

4.3.2.1.4 Control Function 4 - Final Disposal

Techniques considered in this function are similar to those discussed under Section 4.3.1.1.7 for Water Source Type 1 controls. The performances are expected to be similar whether the techniques are applied to Water Source Type 1 or Type 2 waste streams, although pretreatment requirements are different. The costs and performances for these techniques have been presented in Section 4.3.1.1.7 and the appendices.

Water Source Type 2
Coal Pile Runoff
Control Function 1, 2 and 3

4.3.2.2 Stream 012 Coal Pile Runoff

As discussed in Section 3 this stream is not unique to the EDS plant. It is expected that this stream would be acidic and may contain some trace metals leached out from the coal. This is an intermittant stream; its flow rate and characteristics would depend on precipitations (rainfall and snow) and coal characteristics. This stream is typically collected in a sedimentation unit (Function 1 control) where the pH is adjusted (Function 2 control) before discharge. Where zero discharge is desired, some form of volume reduction (Function 3) would be applied. These control techniques have been discussed in Section 4.3.2.1. This stream may be combined with the Combined Inorganic Stream for treatment.

4.4 SOLID WASTE MANAGEMENT

Many solid waste streams are generated in an EDS facility. The characteristics of the streams which are generated from the base plant were presented in Section 3; the streams which result from the pollution control measures, i.e., secondary streams, are discussed in various sections throughout Section 4. Secondary waste streams are fairly significant streams. Depending on the pollution control processes used in the air and water media, they may account for more than 20% of the total solid waste streams generated.

The solid waste streams can be grouped into four waste type categories according to their characteristics and potential control requirements. These waste types are:

- Source Type 1 - Inorganic Ashes and Sludges
- Source Type 2 - Recovered unsalable By-Products
- Source Type 3 - Organic Sludges
- Source Type 4 - Spent Catalysts

Several control techniques are potentially applicable to these wastes. In general, solid waste control techniques aim at containing the entire waste stream. Thus, the performance of these techniques, in terms of removal or control efficiencies, are generally 100 percent. However, unless designed and operated properly, secondary waste streams with undesirable characteristics may be generated and migrated away from the site employing the technique. For example, runoff can contaminate surface water and percolating water can contaminate groundwater. The significance of this depends upon the nature of the species which might be leached out of the solids by the runoff/percolation. Thus, in selecting solid waste management techniques, the major evaluation criteria are whether the technique is applicable and economically feasible, and whether the secondary waste streams are suitably contained.

Based upon current techniques practiced in the synfuel and other industries, together with those being considered by proposed synfuel plants, the bulk of the solid waste from EDS facilities will likely be disposed of on land. Land-based disposal techniques are by far the most site specific techniques. The suitability of the the site, as well as the design and operation of the facility would depend on the site location, transportation costs, hydrogeologic conditions, and many other factors. In short, a detailed analysis of the specific site is an important element of the overall control technique evaluation process.

Land disposal (e.g., landfill, surface impoundment, land treatment) will be subject to regulations promulgated by EPA pursuant to the Resource Conservation and Recovery Act (RCRA), covering the generation, transport, treatment, storage and disposal of solid wastes. Requirements concerning a solid waste can vary signficantly, depending upon whether the waste is determined to be "hazardous" or "non-hazardous" as defined by RCRA regulations. In this section, no attempt is made to judge whether the various individual waste streams will be determined to be hazardous or not. Rather, treatment and disposal techniques are presented which would cover the range of possibilities, whether the waste is hazardous or non-hazardous.

Another technique potentially applicable to some solid waste streams - in addition to land disposal - is incineration. If a waste which is determined to be "hazardous" is proposed for incineration, the incinerator will have to be designed and operated in accordance with RCRA requirements.

The available techniques that may apply to the EDS solid waste streams are identified and evaluated in this section. Since no specific site is being considered, a general overview of these techniques is first presented. This is followed by an evaluation of the applicable controls to each individual stream under each waste source type. The emphasis is on identifying the applicability of the technique. For the reasons stated above and because the characteristics of many of the solid waste streams are not known, it is not possible to evaluate the optimum design and operation of these techniques in the

PCTM; optimum design/operation will vary with the site. It is assumed that solid waste management facilities are captive, i.e., they only handle waste from the EDS plant. Those situations in which solid waste streams can be combined for treatment or disposal are pointed out.

The organization of this section diverges slightly from the organization of the air and water media sections. The applicable control functions will be presented first because these functions generally apply to all of the waste types. This is followed by an evaluation of the applicable controls to each waste stream under each waste source type. Some examples illustrating how streams might be combined for disposal in an integrated plant, and some examples of how techniques may be combined to control specific streams, are presented in the last part of this section.

4.4.1 Solid Waste Control Functions

Several control techniques are potentially applicable for managing the solid waste streams. These techniques are summarized in Table 4-78. As shown, they can be broadly divided into three control categories according to their functions. The three control functions are:

- Control Function 1 - Reuse/Resource Recovery
- Control Function 2 - Treatment
- Control Function 3 - Disposal

Treatment may involve specific chemical/physical processes for preparing the waste to meet certain reuse/resource recovery specifications or to stabilize the waste for disposal. Reuse/resource recovery is one form of ultimate control for the waste. This approach is usually waste specific, highly dependent on market availability and cost tradeoffs, and may require specific treatment of the waste. Disposal is another form of ultimate control for the waste. Most disposal techniques are land-based techniques and thus are highly site-specific. The major site-specific factors that affect the design, operation and cost of land-based techniques are summarized in Table 4-79. The following is a brief description of the individual techniques. A more detailed discussion is included in Appendix C.

4.4.1.1 Control Function 1 - Reuse/Resource Recovery

Reuse or resource recovery of waste streams is desirable from environmental standpoint because less wastes require disposal (if any) and other resource requirements are reduced. Potential adverse environmental impacts associated with disposal of the waste are eliminated although other impacts may arise as a result of the reuse/recovery process utilized. This control approach is highly waste specific and is constrained by the availability of markets or uses for the waste. Available reuse/resource recovery alternatives will be identified under each specific waste stream discussion.

TABLE 4-78. SUMMARY OF SOLID WASTE MANAGEMENT TECHNOLOGIES

Technology	Description	Operations Considerations	Reliability or Limitations	Equipment	Waste Generated	General Comments
Reuse/Resource Recovery						
Reuse, Resource Recovery	Waste is utilized in a manufacturing process, construction application, or is processed for reuse in original application, or valuable components are recovered from waste as byproduct for marketing.	User for waste must be located; cessation of reuse requires immediate alteration of management techniques; thereby, necessitating long term contracts	Markets for wastes are limited and economic viability is heavily influenced by distance to market	Transportation vehicles	None	In general, these are the most environmentally acceptable management techniques.
Treatment						
Fixation/ Encapsulation	Chemicals are added to stabilize or solidify the wastes	Wide variety of wastes can be processed; the feasibility of solidifying a particular waste may differ with different process	Limited commercial experiences	May include mixing chamber, pumps, metering devices, storage tanks, chemical feed systems	Solidified waste	Most processes are applicable only to small waste streams
Incineration	Organic Wastes undergo destruction to reduce volume and toxicity	Each type of organic waste may require different operating characteristics; experienced operators required	Only organic wastes can be processed	Waste transportation machinery, incinerator	Ash; air pollution control residues	Process is energy intensive
Disposal						
Landfill	Site is designed, constructed, and operated to totally contain waste; features include natural or artificial liners, leachate collection and treatment systems, and groundwater monitoring system	Wide variety of wastes can be accepted; provisions must be made for incompatible wastes and reactive wastes	Waste will be contained subject to adequate site operation and maintenance and absence of extreme hydrogeological changes or earth movements	Earth moving equipment, waste handling machinery such as forklifts and trucks	Surface runoff and leachate	Site location and design dependent upon hydrogeological conditions; provisions must be made for site care after cessation of operations
Surface Impoundment	Site is excavated or diked to form pond to contain waste; supernatant is syphoned off and treated or allowed to evaporate	Similar to landfill	Similar to landfill	Machinery to move waste to site, usually pipelines	Supernatant and leachate	Waste may have to be removed when operation ceases; if waste remains, long term site care and maintenance program must be established
Land Treatment	Waste is treated by incorporation into the land according to specific procedures	Only limited types and mass of organic wastes can be managed	Heavily dependent upon weather conditions	Waste moving machinery, usually trucks, and waste incorporation machinery, usually bulldozers and discs	Possibly surface runoff and leachate	Site location dependent on soil conditions; provisions must be made for long term site care
Deep Well Injection	Wastes are pumped through wells into appropriate formations generally several thousand feet below the surface	Wastes frequently must be treated before injection	Injection rates frequently limited by receiving formation	Pumps, injection wells	Residues from the waste treatment process	Some states prohibit deep well injection

TABLE 4-79. SITE-SPECIFIC FACTORS TO BE CONSIDERED FOR LAND DISPOSAL OPTIONS

Climatological

- Wind conditions (e.g., speed, directional flux, dilution factors, humidity, temperature, etc.)
- Precipitation (e.g., annual precipitation, storm intensity, snow contributions, etc.)
- Evapotranspiration rate (e.g., season variations, etc.)

Geologic Factors

- Physiographic features (e.g., runoff coefficient, slope, drainage patterns - dendritic trellis, annular, etc., erosional features, etc.)
- Surface and subsurface geology (e.g., outcrops, bedrock features, strike and dip of the bedrock, rock composition, etc.)
- Soil types (e.g., CEC capacity, texture, permeability, stratification, homogenous vs. heterogenous deposition, chemical composition, percent of humic material, etc.)
- Seismic factors (e.g., ground shaking or rupture)

Hydrogeologic Factors

- Drainage patterns
- Stream flow (e.g., velocity, perennial vs. intermittent, effluent or influent source, etc.)
- Surface waters (e.g., tidal effects, recharge vs. discharge points, etc.)
- Vadose zone (e.g., depth, moisture content, hysteresis patterns, storage capacity, etc.)
- Groundwater (e.g., depth, number of aquifers and relationships, confined or artesian, nature of confining layer(s), capillary fringe characteristics, etc.)
- Piezometric surface (e.g., streamline flux patterns due to seasonal or event related phenomena, influence of recharge/discharge areas, streamline anomalies, etc.)
- Water quality (e.g., background vs. undersite vs. downgradient, water uses - consumptive, irrigation, recreation, point source contributors and their respective hydrogeologic pathways, etc.)
- Floodplain (100 year flood) (e.g., aerial flooding limits, degree of localized streamline pattern reversal, erosional consequences, etc.)
- Wetlands (e.g., recharge vs. discharge source, wetland/groundwater continuity and pathway, etc.)
- Recharge and discharge areas (e.g., proximity of disposal area, volume of flow, etc.)

Land Use Factors

- Historic significance
- Transportation corridor (access)
- Beneficial uses
- Cost
- Demographic setting
- Geopolitical impact
- Ultimate land use

The economics of reuse/resource recovery is sensitive to site-specific factors such as transportation costs and some general factors such as the prices of the recovered/reusable materials and the cost of preparing the waste for reuse/resource recovery. The feasibility of this control should be thoroughly analyzed for each individual facility before implementation.

4.4.1.2 Control Function 2 - Treatment

For the purpose of this manual, only two solid waste treatment techniques are considered. These are fixation/encapsulation, and incineration. Other techniques such as dewatering, neutralization, etc. are either considered as part of the base plant operation, or part of the pollution control processes in other media (air or water), depending on the origin of the waste.

4.4.1.2.1 Control Technique 1 - Fixation/Encapsulation

Fixation and encapsulation are treatment processes which stabilize or solidify waste constituents, or enclose the waste within other materials. Fixation processes generally combine the concepts of solidification (the alteration of the characteristics of a waste to attain desired structural characteristics) and stabilization (the immobilization of waste constituents by chemical reactions to form insoluble compounds or by entrapment within an inert polymer or stable crystal lattice). Depending on the principal chemical agents used, fixation processes can be categorized as cement-based, lime-based, thermoplastic organic polymer, and glassification techniques. Encapsulation processes involve enclosing the waste in a coating or jacket of an inert, relatively impermeable material so that contact between the waste and water is prevented. Regardless of the specific chemicals used, typical fixation process operations involve mixing the chemical with the waste in a reactor at specific temperature and for a specific time period. The end product is the fixed waste.

In principle, these processes are applicable to treating any waste by applying "sufficient" chemicals. For economic reasons, these techniques have only been applied to small volume waste streams or streams which are prone to pozzolanic reactions. Chemical requirements for fixing the latter type of streams are generally low. FGD sludge is one example of this type of stream. Several proprietary, cement- or lime-based fixation techniques have been used in fixing FGD sludges. Typical total annualized unit fixation costs, not including ultimate disposal costs, are reported to be \$10 to \$17 per metric ton of sludge fixed (100-103). It is recommended that, before implementing this technique to a specific waste, detailed treatability studies with various chemical additives be performed to: 1) establish that the waste is treatable, 2) select the optimal process, and 3) minimize the cost (100).

4.4.1.2.2 Control Technique 2 - Incineration

Incineration is a controlled thermal decomposition process which reduces the weight, volume and characteristics of the waste by oxidizing the organic compounds present in the waste. The extent of volume and weight reduction is dependent upon the waste characteristics, the incineration process, and the incinerator operating condition. The major concerns in incineration are getting adequate destruction of the organics and of avoiding air emission of any unburned organics. The end products of incineration include carbon dioxide, water, ash, and other inorganic compounds. Incineration has been applied to various industrial wastes including refinery wastes, sewage sludge, paper mill waste liquor, pharmaceutical wastes, and organic chemical wastes. The common types of incinerators used for solid waste disposal include rotary kiln, multiple hearth, and fluid bed reactor (104). The capital investment for a 9.2×10^6 Btu/hr rotary kiln incinerator with energy recovery is estimated to be 16 to 20 million dollars (see Figure C6-2 in Appendix C6). The total annualized cost for this is estimated to be 9 million dollars per year (see Figure C6-4).

4.4.1.3 Control Function 3 - Disposal

As mentioned before the bulk of the solid wastes are likely disposed on land. Three potentially applicable techniques are discussed in this section. These are landfill, surface impoundment, and land treatment.

4.4.1.3.1 Control Technique 1 - Landfills

Landfills have been widely used for the disposal of municipal refuse and a range of industrial wastes. In landfilling, waste is brought to the disposal site by truck or conveyor, spread in layers, and compacted with heavy equipment. In most landfills the waste is covered with a thin layer of soil at the end of the working day. The process is repeated until the desired depth is reached or the available area is filled. A final cover of soil is then added. The finished site is either revegetated or prepared for other end uses.

Landfills can be accomplished in unexcavated depressions (the area-fill method) or in excavated sites (the trench-fill method). These can be natural sites or man-made sites such as coal mines. There are two major concerns in landfill design and operation. Runoff from landfill sites may contaminate surface water; and percolation from sites, after passing through the waste pile, may contaminate groundwater. Runoff/surface water contamination may be prevented by grading of the site, and containment of runoff. Diversion channels should be incorporated into the initial design of the landfill and constructed before the site begins accepting waste (105). This prevents surface runoff water from the surrounding terrain from entering the site and generating leachate.

Migration of leachate into the groundwater can be controlled by lining the landfill with clay, concrete, asphalt or plastic. Liners will often be required if the solid waste is considered to be hazardous; liners may sometimes also be desirable if the waste is non-hazardous. The choice of an appropriate liner or liners will depend on site-specific climatologic, geologic, and hydrogeologic factors, as well as on the compatibility of the liner and the waste to be contained and the relative cost of compatible liners. A leachate collection and treatment system may also be necessary. Such systems consist of perforated pipes and sumps placed in a layer of permeable sand at the bottom of the fill. After being pumped out of the landfill, the collected leachate may be treated in the EDS facility's wastewater treatment system or in a separate treatment system (106).

In the absence of any judgement concerning whether a given waste might ultimately be determined to be hazardous or not, and in order to remain independent of site-specific factors, two landfill designs are considered in the PCTM. These two designs cover the range from the simplest set of conditions (non-hazardous waste with favorable hydrogeologic and other site conditions that preclude needs of liners) to the most complex (hazardous waste with unfavorable hydrogeologic and other site conditions that a double liner system is felt necessary). The two landfill designs are presented in Figure 4-7. For the purposes of this assessment, the lined landfill design assumes an upper liner consisting of 1 m of clay, and a lower liner of 0.076 cm (30 mil) synthetic material. Both the unlined and the lined landfill designs assume the completed fill will be 30 meters above the original land surface with a slope of 3:1. Both landfills will have a final cover consisting of 0.5 m of sand and 0.3 m of clay. The most complex hazardous case would also include special provisions for closure and post-closure care, monitoring, record keeping and other requirements.

The capital investment and annualized unit costs as a function of the site capacities for the two designs are presented in Figure 4-8. To be consistent with cost estimates performed for the base plant and the air and water

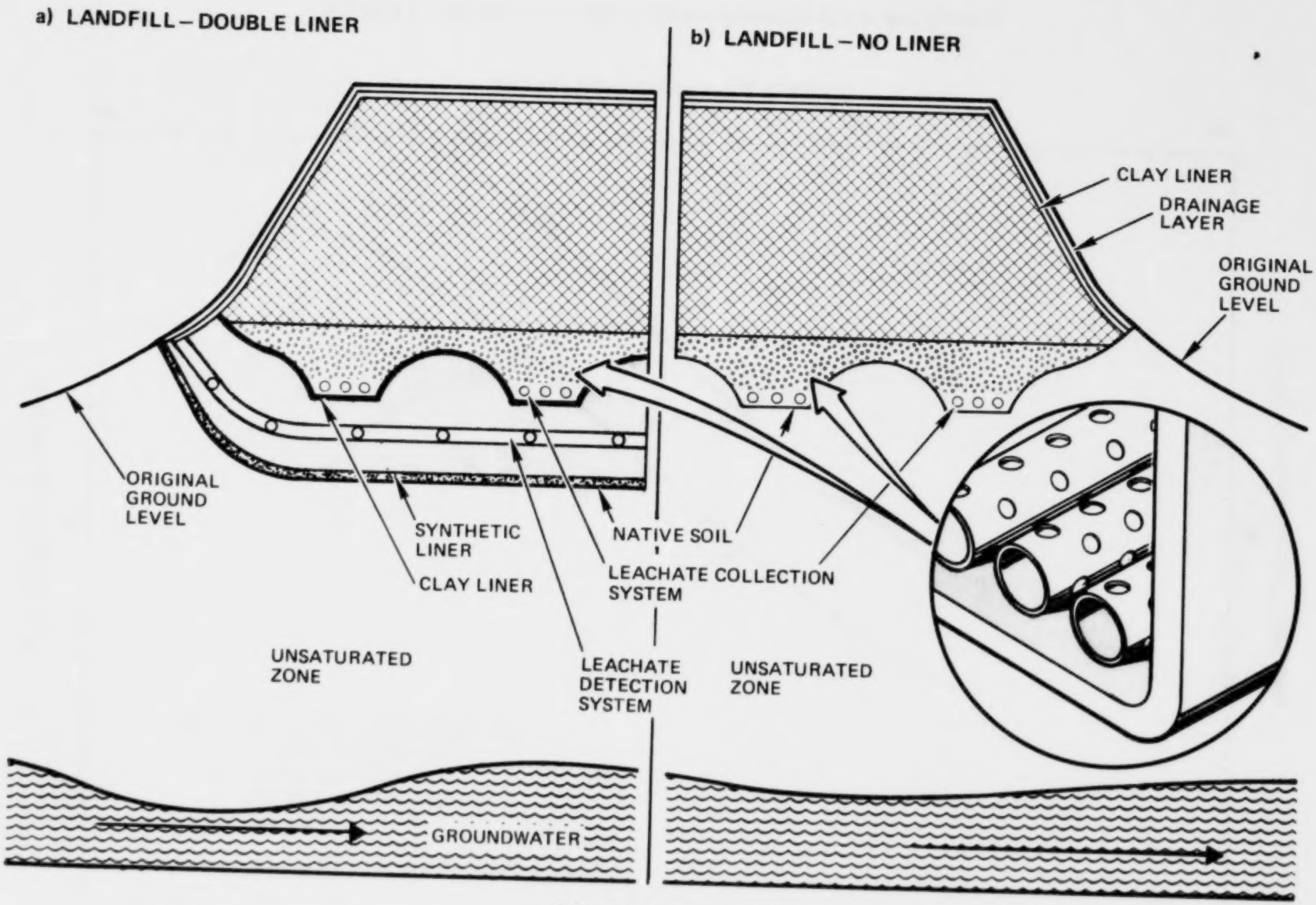
a) LANDFILL – DOUBLE LINER
b) LANDFILL – NO LINER
CLAY LINER
DRAINAGE LAYER
ORIGINAL GROUND LEVEL
ORIGINAL GROUND LEVEL
SYNTHETIC LINER
CLAY LINER
NATIVE SOIL
LEACHATE COLLECTION SYSTEM
LEACHATE DETECTION SYSTEM
UNSATURATED ZONE
UNSATURATED ZONE
GROUNDWATER

Figure 4-7. Landfill design.

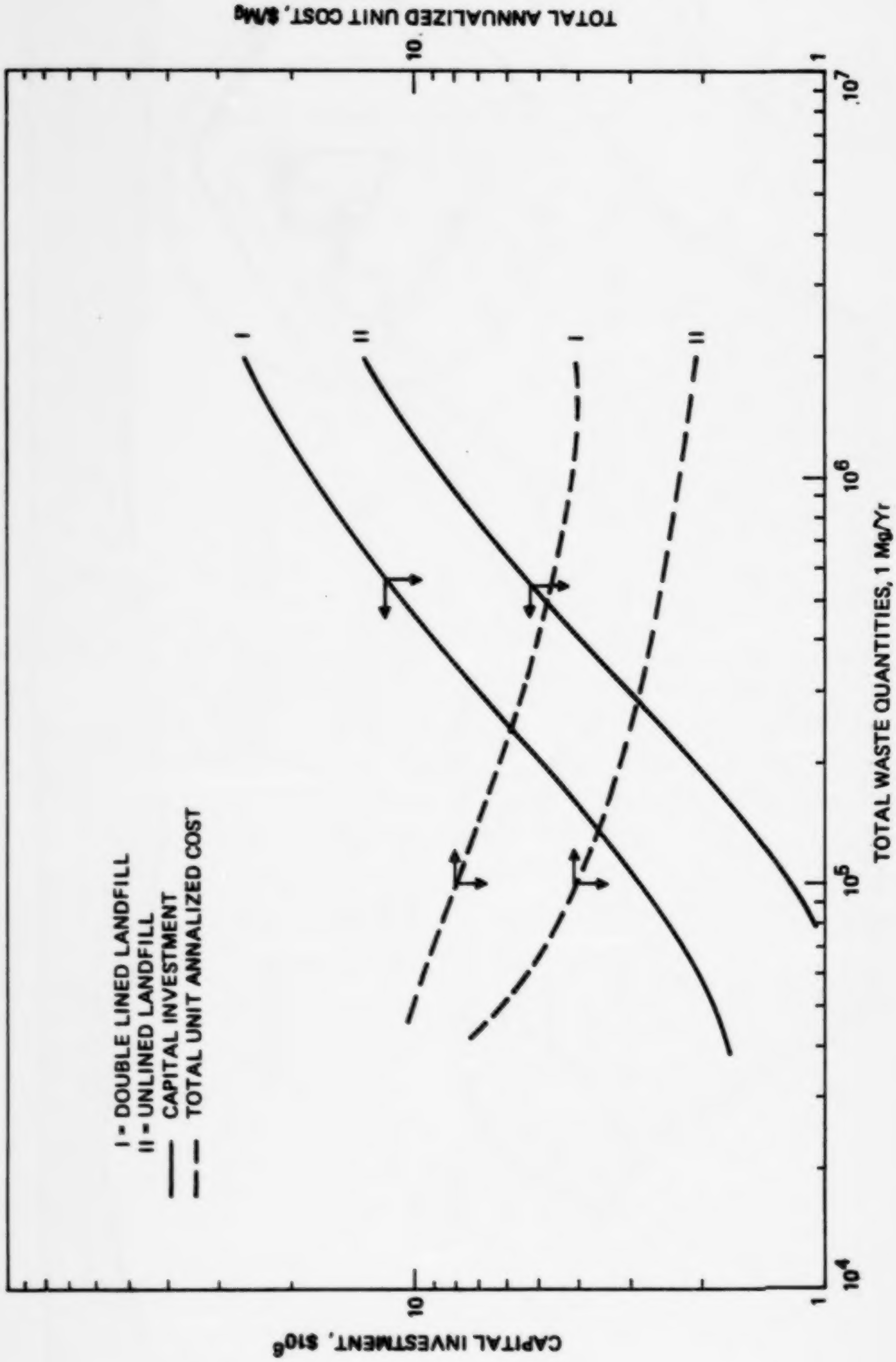

Figure 4-8. Estimated Landfill Costs as a Function of Landfill Capacity

media, no land cost is included.* The capital investment developed includes site preparation cost (clearing and scrubbing, groundwater monitoring and collection systems, liners), final cover and revegetation cost, and landfill equipment cost (see Appendix C-2, section 7 for unit installed equipment costs for these items and Table 4-1 for bases for converting installed equipment costs into capital investment). It was assumed that no excavation is required. The annualized unit cost includes labor, fuel, leachate collection and analysis and amortized capital investment; but does not include hauling cost and other costs such as administrative, closure/post-closure and liability costs. These other costs would depend on the classification of wastes under RCRA. EPA has estimated that for a 50,000 metric ton per year commercial hazardous waste landfill, administrative and other compliance costs amount to \$9/Mg (107). Hauling cost is a function of distance between the plant and the disposal site. It is estimated that the unit cost for a round-trip distance of 5 and 15 km are \$2 and \$4/Mg, respectively.

4.4.1.3.2 Control Technique 2 - Surface Impoundments

Surface impoundments have been utilized widely by municipalities and industries to process or dispose of waste liquids, sludges, and slurries. Like landfill sites, the impoundments can be in natural depressions or in excavated areas. Earthen dikes are usually constructed around the impoundment area. Wastes are usually transported hydraulically to the impoundment. The wastes deposit at the bottom of the impoundment; the supernatant may be removed and treated for discharge or recycle, or allowed to evaporate.

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* About 1,100,000 m^2 of land is required per 10^6 Mg/yr of waste generated, or about 110 hectare/10^6 Mg/yr (130 acre/10^6 ton/yr). Assuming a land cost of \$5,000/10,000 m^2 the capital investments for the lined and unlined landfills, as presented in Figure 4.4.2, will be increased by 9 and 12 percent, respectively. Additional land may be required for road construction, buffer zone, buildings, etc., which will incur additional costs.

Leachate migration from surface impoundments is controlled in much the same way leachate is controlled from landfills. Diversion structures prevent runoff from surrounding terrain from entering the site; liners of in-place or compacted soils or synthetic materials retard leachate migration down into the soil and groundwater. As in the case of landfills, the PCTM considers both an unlined and a double-lined impoundment, in order to cover the range of possible experience. The unlined impoundment represents the simplest set of conditions (non-hazardous waste with favorable hydrogeologic and other site conditions that preclude the needs for liners), and the lined impoundment represents a much more complex situation (hazardous waste with unfavorable hydrogeologic and other site conditions that a double liner system is felt necessary).

When the surface impoundment has been filled with waste, the site may be closed in one of two ways: the residual solids may be left in place and covered with clay and sand to prevent erosion and the infiltration of precipitation, or the solids may be removed from the impoundment site for further treatment or final disposal in a landfill. If wastes are left in place, the site becomes a landfill (subject to any requirements pertaining to a landfill), and a long-term site care and maintenance program will need to be established. The cost per unit volume for surface impoundments are expected to be similar to those of landfills with similar depths. However, since surface impoundments generally are used for the disposal of wet, not yet dewatered wastes, a larger area may be required per mass of dry solid, resulting in higher disposal costs, i.e., some of the disposal cost will be for 'disposing' of water. Disposal cost could be reduced if the wet wastes were dewatered first but there would be costs associated with dewatering. The trade-off in disposal cost versus dewatering cost is highly dependent upon waste characteristics and site-specific factors.

4.4.1.3.3 Control Technique 3 - Land Treatment

Land treatment refers to the use of soil as a medium to treat and dispose of waste. Also known as landfarming, landspreading, and soil application, land treatment has been practiced successfully for the treatment and disposal of municipal wastewater treatment sludges and petroleum industry oily wastes for many years. It relies on the ability of naturally occurring soil microorganisms to decompose and utilize organic compounds under aerobic conditions. The design and operation of land treatment systems would be affected by whether or not the wastes were considered to be hazardous.

Wastes added to soil are subject to one or more of the following processes: (1) decomposition/degradation; (2) leaching; (3) volatilization; and (4) incorporation into the soil matrix (e.g., through ion-exchange or adsorption). It is the degradation processes which treat the waste to reduce its objectionable (e.g., toxic) properties; these processes must be maximized during land treatment, while the other processes must be minimized or eliminated. Applying biodegradable wastes, maintaining proper (aerobic) conditions for microbial action, and avoiding or pretreating wastes which are toxic to the microorganisms will encourage degradation processes. Proper site selection and proper site management will minimize leaching and subsequent contamination of surface water and groundwater. If volatile wastes are to be land-treated, subsurface injection of the wastes or immediate tilling after application will minimize air pollution.

Wastes with high concentrations of toxic substances such as arsenic, cadmium, lead, and mercury should not be land treated in sites where food chain crops are grown, as they may be incorporated into the soil and taken up by plants (108). Prior to land treating biological oxidation sludge, long-term studies should be performed to confirm that the waste is degradable in the soil and that there is no accumulation of non-degradable toxic substances in the soil.

Solid Waste
Control Function 3
Disposal

Assuming biosludge is applied on land 10 times/yr, with an application rate of 60 Mg/acre/application and a factor of 2 to account for land required for roads, buffer zones, dikes, etc., it is estimated that 430,000 m^2 of land is required to land-treat 100 Mg/day of biosludge. The capital investment for this site would be 0.76 million dollars. This cost was calculated from installed equipment cost using the procedures summarized in Table 4-1. The installed equipment costs include land preparation cost ($0.52/$m^2$), waste spreading equipment costs ($160,000) and monitoring well costs ($25,000); but not land cost. Detailed bases and assumptions for the design of and cost estimates for land treatment are presented in Appendix C-4.

4.4.2 Source Type 1 - Inorganic Ashes and Sludges

Inorganic ashes and sludges are the largest volume solid waste type from an EDS facility. This source type includes 12 streams for the base case and 13 streams for the MFS case. The flow rates of these streams are summarized in Table 4-80, and the application of the available techniques to these streams is evaluated in this section.

TABLE 4-80. SUMMARY OF SOLID WASTE QUANTITIES FOR SOURCE TYPE 1 STREAMS

Stream No.	Stream Description	Quantity, Mg/yr	
		Base Case	MFS Case
(302)	Flexicoking Gasifier/Heater Dry Fines	138,000	66,000
(303)	Flexicoking Gasifier/Heater Wet Fines	345,000	164,000
(306)	Flexicoking Gasifier/Heater Bed Coke	707,000	363,000
(313)	Flexicoking Heater/Reactor Chunks/Agglomerate	20,000	10,000
(108)	Solids Accumulated in Slurry Drier	820	820
(704)	Bottom Ash from Steam Generation System	3,900	10,000
(710)	Bottom Ash from Power Generation System	12,600	8,000
(703)	Fly Ash from Steam Generation System	17,000	44,000
(709)	Fly Ash from Power Generation System	46,000	33,000
(705)	FGD Sludge from Steam Generation System	51,000	150,000
(711)	FGD Sludge from Power Generation System	162,000	116,000
(722)	Raw Water Treatment Sludge	14,000	15,000
(442)	Slag from Partial Oxidation Unit	0	447,000
	Total	1,520,000	1,430,000

4.4.2.1 Combined Flexicoking Waste Stream - Streams 302, 303, 306

As presented in Section 3.3.4.1.3 these streams contain fairly high levels of carbon (10 to 40%) with most of the remainder being coal ash. Leachate may include low levels of trace metals (see Table 3-39 and 3-40 for laboratory simulated leachate characteristics); tests of leachate for organics have not been conducted, but organics levels would be expected to be low. Although stream 303 is wet, containing about 40% water and the other two are dry, they are likely combined for treatment and disposal. As shown in Table 4-80, these three streams total 1.19×10^6 Mg/yr in the base case and 0.59×10^6 Mg/yr in the MFS case.

4.4.2.1.1 Control Function 1 - Resource Recovery

Due to its high carbon content this stream may be used as fuel in a boiler. However this material may present two major problems with respect to conventional pulverized coal-fired boilers. First, this material contains little or no volatiles which may result in flame stability problem unless a readily combustible supplemental fuel is added or other countermeasures are taken. Second, the high moisture content in Stream 303 may require extensive drying and thus special handling and drying equipment. For these reasons, fluidized bed combustion boiler (FBC) is used as an example to illustrate this resource recovery alternative. The use of FBC boiler is being considered for at least one proposed Koppers-Totzek gasification facility in the U.S. (K-T gasifier generates a dust which may contain up to 50% water, 20% carbon and 30% ash, not unlike the Flexicoking solids).

4.4.2.1.1.1 Control Technique 1 - FBC Boiler

An FBC boiler is comprised of a granular bed material which is suspended or "fluidized" by a stream of air. The fuel is injected into this bed and burned. Alkaline sorbents, typically limestone or dolomite, may also be injected into the bed to react with SO_2 formed during combustion. The inert material of the fuel usually leaves the top part of the bed with the flue gas, and the spent sorbent is removed from the bottom of the bed.

Two secondary waste streams are generated from the FBC boiler, namely a flue gas stream and a spent sorbent stream. The flue gas is expected to contain some NO_x, SO_2 and particulates. NO_x control is inherent in FBC; any SO_2 control is done by bed sorbent. Thus, only particulate control is of concern. The applicable controls for the flue gas have been discussed in Section 4.2. Because of subsequent particulate emission controls applied, essentially all of the ash in the flue gas will be captured and will appear as a solid waste. The quantities of the ash generated for the base case and the MFS case are estimated to be 6.5×10^5 Mg/yr and 3.0×10^5 Mg/yr, respectively. The characteristics of this stream are expected to be similar to the fly ash from the conventional coal-fired boiler. The quantities of spent bed materials are estimated to be 6,600 Mg/yr and 3,300 Mg/yr, respectively, for the two cases (109). This material contains mainly $CaSO_4$, CaO, $CaCO_3$ and some inerts.

Assuming a carbon heating value of 850 Kcal/Kg (1554 Btu/lb), heat input to the FBC boiler from the base case and MFS case will be 38×10^6 kcal/hr (150×10^6 Btu/hr) and 19×10^6 kcal/hr (75×10^6 Btu/hr), respectively (see Table 3-78 and 3-79 for characteristics of the individual streams). The capital investments for these boilers are estimated to be 11 and 6.5 million dollars (see Table 4-81). These costs are derived at using the procedures presented in Table 4-1 and the installed equipment costs reported in Table A-4 of reference 109, updated to 1980 bases. The installed equipment cost reported in reference 109 are for FBC boilers burning high sulfur coals; these costs may be low for boilers burning Flexicoking wastes. Assuming a labor requirement of 2 operators per shift, no steam credits and using the bases summarized in Table 4-2, the total annualized costs for the base case and MFS case are estimated to be 2.5 and 1.6 million dollars per year, or about 0.26% and 0.19% of the uncontrolled base plant costs.

TABLE 4- 81. SUMMARY OF ESTIMATED COSTS FOR FBC BOILERS APPLIED TO COMBINED FLEXICOKING SOLIDS (1,134 Mg/hr Feed Coal EDS Plant)

Item	Base Case	MFS Case
Capital Investment, $\$10^6$	11	6.5
Total Annualized Cost, $\$10^6$	2.5	1.6
Unit Capital Investment, $/Mg/hr*	73,000	86,000
Unit Annualized Cost, $/Mg*	2.1	2.7
% Base Plant Capital Investment, %	0.34	0.23
% Base Plant Annualized Cost, %	0.26	0.19

*Expressed as Mg of Flexicoking solids burned.

Solid Waste Source Type 1
Combined Flexicoking Waste Streams
Control Function 1 -
Resource Recovery

It is estimated that 4.1×10^5 Mg/yr and 2.1×10^5 Mg/yr of steam is generated by the two cases. A steam credit of \$6.1/Mg and \$7.6/Mg would be necessary for the base case and MFS case respectively, if the FBC boiler cost were to break even, i.e., if the steam credit were to offset the capitalized and operating costs. This shows that the cost estimates for burning this stream in FBC boilers is very sensitive to the steam credit.

One other factor that affects the economics of this alternative is the total amount of waste generated that requires disposal. Burning this stream will reduce its quantity by more than 40% (from 1.19×10^6 Mg/yr of Flexicoker wastes to 5.1×10^5 Mg/yr of dry FBC boiler ash and spent bed media for the base case; from 5.9×10^5 Mg/yr to 3×10^5 Mg/yr for the MFS case). Recovering the steam for in-plant use will result in less steam being required from the separate coal-fired steam boiler. The net result is that less ash is being generated from the steam/power generation part of the plant. It is estimated that for the base case and MFS case, respectively, about 6 Mg/hr of coal feed to the coal-fired boiler would be displaced by the burning of the Flexicoking waste. These amount to reductions of 16% and 3% of the coal boiler size. The associated reduction in boiler ashes and sludges are estimated to be 12,000 Mg/yr and 6,000 Mg/yr (see Table 4-80). Thus, in effect the use of FBC boiler for burning the Flexicoker wastes will reduce the total solid waste quantities that require ultimate disposal by 6.92×10^5 Mg/yr and 2.84×10^5 Mg/yr for the base case and MFS case, respectively. Assuming annualized unit costs of \$2.5 and \$4.1/Mg for a nonlined and lined landfill, this reduction in waste generation would amount to savings of 1.7 and 2.8 million dollars per year for the base case, and 0.76 and 1.2 million dollars for the MFS case.

4.4.2.1.2 Control Function 2 - Treatment

4.4.2.1.2.1 Control Technique 1 - Fixation

The techniques that are potentially applicable to treating this stream are fixation processes. Fixation of this stream may be appropriate if data indicate that significant concentration of trace metals or unexpected high levels of organics are detected in leachates generated from this material. The potential need for this treatment has not been established at the present time; available data indicate trace element levels in leachates from the Flexicoking waste are low.

The performance and cost of fixing this stream is dependent upon the specific process (additive) used, which can only be established after thorough treatability studies. Lime- and cement-based fixation techniques, which have been applied to boiler fly ash, FGD sludges and coal fines are likely candidates for treating this stream. From available data on the characteristics of this stream, it is entirely possible that much more chemical will be needed to fix this stream than to fix an equal amount of FGD sludge. Thus, the unit cost may be much higher than the $10 - $17 per Mg reported for FGD applications (100-103).

4.2.2.1.3 Control Function 3 - Disposal

Techniques potentially applicable to disposing this stream include landfill and surface impoundment.

4.4.2.1.3.1 Control Technique 1 - Landfill

In landfilling, the Flexicoking wastes are likely transported to the disposal site by truck, spread on the surface of land or previously placed wastes, and compacted. As the pile height increases, a working face with safety slope is developed to ensure stability of the fill.

If the Flexicoking wastes are determined to be non-hazardous, then, in the most favorable case (e.g., favorable site conditions), an unlined landfill might be possible. On the other hand, if the wastes were considered to contain hazardous components, a lined landfill would be necessary where hydrogeologic or other site factors are unfavorable.

Solid Waste Source Type 3
Combined Flexicoking Waste Stream
Control Function 3 -
Disposal

Based on current and proposed practices in the synfuel and other industries, this stream is likely co-disposed with some other solid waste streams from the plant in one common landfill. Although more than one landfill/disposal facility (e.g., one landfill design for hazardous waste and one landfill for non-hazardous waste) may be operated in an EDS plant, for costing purposes in the PCTM one landfill accepting all the wastes from the EDS plant is assumed. By considering the alternatives of all wastes being disposed of in a non-hazardous waste landfill with no liners and a hazardous waste landfill with double liners, the range of landfill costs in the PCTM should bracket the costs that might be encountered in practice for any split of the wastes between hazardous and non-hazardous categories.

As summarized in Table 4-80 the total solid waste quantities generated from the base case and MFS case are, respectively, 1.6 and 1.5 million Mg/yr. The capital investments for landfills accepting these quantities of wastes are estimated to be the same - 11 million dollars for a non-lined landfill and 23 million dollars for a lined landfill (see Figure 4-7 for landfill design and Figure 4-8 for estimated capital investments). The unit annualized costs for the non-lined and lined landfill will be \$2.3/Mg and \$4.1/Mg, respectively (see Figure 4-8). Thus, assuming the Combined Flexicoking Wastes Stream are disposed of in one common landfill that accepts all other wastes from the EDS facility, the total annualized cost that would be attributable to the Combined Flexicoking Waste Stream alone would be \$2.7 million/yr (for disposal in a non-lined landfill) and \$4.9 million/yr (for disposal in a lined landfill) for the base case; and \$1.4 million/yr and \$2.4 million/yr for the MFS case.

If the Combined Flexicoking Waste Stream were disposed of by itself in a separate, dedicated landfill, costs for this smaller landfill (1.2×10^6 Mg/yr for base case, 0.59×10^6 Mg/yr for MFS case) would be almost as great as for a total combined landfill handling all wastes. The capital investment for unlined and lined landfill accepting Combined Flexicoking Waste Stream alone would be \$10 and \$20 million for the base case, respectively; and \$0.4 and \$10 million for the MFS case (see Figure 4-8).

4.4.2.1.3.2 Control Technique 2 - Surface Impoundment

In surface impoundment, the Flexicoking wastes would be transported hydraulically to the site. The wastes would settle to the bottom of the impoundment, the supernatent being recycled or discharged. The disposal cost per unit volume for surface impoundment are expected to be similar to landfills with similar depths. The total disposal cost may be higher because surface impoundments generally are used for the disposal of wet, not yet dewatered wastes; extra capacities are needed for the water content in the waste. This cost may be partially or totally offset by the reduction in costs associated with dewatering the wastes.

Solid Waste Source Type 1
Flexicoking Heater/Reactor
Chucks/Agglomerate
Control Functions 2 and 3

4.4.2.2 Stream 313 - Flexicoking Heater/Reactor Chunks/Agglomerate

This stream contains large chunks of inert material originating from the feed coal. Its characteristics are expected to be similar to boiler bottom ash. Applicable control techniques include fixation and landfill. As with the Combined Flexicoking Wastes the need for fixing this stream has not been established. Where applicable, the likely fixation process candidates for fixing this stream are lime- and cement-based techniques. From available data on the characteristics of this stream (mainly on the lime content) it is entirely possible that much more chemical will be needed to fix this stream than to fix an equal amount of FGD sludge. Thus, the unit cost may be much higher than the \$10 - \$17/Mg reported for FGD application (100-103).

In landfilling, this waste is expected to be co-disposed with the Combined Flexicoking Waste Stream and all other solid wastes in a common landfill, as discussed in the previous section. The annualized unit cost of disposing this stream in a common landfill would be \$2.3 and \$4.1/Mg, respectively, for the non-lined and lined landfill. For this stream (20,000 Mg/yr for base case, and 12,000 Mg/yr for MFS case) these unit costs translate into incremental annual costs of \$46,000 and \$82,000/yr for the base case; \$23,000 and \$41,000/yr for the MFS case (not including hauling and administrative costs). Again, the consideration of both a lined and unlined landfill is intended to bracket the range of conditions that might be encountered in any individual situation.

4.4.2.3 Stream 108 - Solids Accumulated in Slurry Drier

This is a relatively small volume solid waste stream (820 Mg/yr) which consists essentially of part feed coal and part recycled solvent. There are no data on the characteristics of leachates generated from this waste. Potentially applicable techniques are discussed below.

4.4.2.3.1 Control Function 1 - Resource Recovery

Due to its high carbon content, this material can be used as feed to the boilers, Flexicoker, partial oxidation unit or the FBC boiler as discussed in 4.4.2.1.1. Pre-sizing of the material may be required before it can be used as feed. This is a small stream with flow rates of about 820 Mg/yr. Feeding this to the above mentioned units is not expected to incur any significant cost or savings.

4.4.2.3.2 Control Function 2 - Treatment

If this stream were not returned to the process (per section 4.4.2.3.1) applicable treatment techniques might include fixation/encapsulation processes. Fixing this stream will reduce its leaching potential. This may be appropriate due to its high organic content.

The performance and cost of fixing this stream is dependent upon the specific process used. The presence of solvent and the high carbon content of this stream will likely interfere with the pozzolanic reactions in the fixation processes. Thus, it is expected that much more chemical will be needed to fix this material than to fix an equal amount of FGD sludge. The unit cost may be much higher than the $10 - $17/Mg reported for fixing FGD sludges (100-103).

4.4.2.3.3 Control Function 3 - Disposal

Applicable disposal techniques include landfill and surface impoundment. If disposed, this stream is assumed to be co-disposed with other Source Type 1 streams in a common landfill (see 4.4.2.1.3.1). The annualized unit cost of disposing this stream in a common landfill would be $2.3 and $4.1/Mg, respectively, for an unlined and lined landfill. For this stream, these unit costs translate into incremental costs of $1,900 and $3,400/yr, respectively.

4.4.2.4 Stream 442 - Slag from Partial Oxidation Unit

The slag from partial oxidation unit, which exists only in the MFS case, is a coarse pebble-sized material which is physically stable and essentially chemically inert. As discussed in Section 3, this stream is essentially coal ash with little or no carbon if the gasifier is operated properly. Leachate may include low levels of trace metals (see Table 3-58 and 3-59 for laboratory simulated leachate characteristics). Tests of leachate for organics have not been conducted, but organics levels would be expected to be low.

4.4.2.4.1 Control Function 1 - Resource Recovery

The slag may be utilized in a number of commercial applications, just as boiler bottom and fly ash from fossil-fueled power plants have been used. The National Ash Association reported that 24.3 percent of the coal boiler ash produced in 1977 was reused in commercial applications (111). Ash has been used commercially as a partial replacement for cement in concrete, as fill material for roads and other construction projects, and as blast grit and roofing granules. The slag may need to be dried, crushed, and sized before it can be used in such applications.

The major constraints on reuse of the slag are market limitations. Market conditions will vary from site to site. Given the fact that not all ashes from existing power plants are commercially utilized, it may be difficult to find markets where all or significant quantities of the slag from the EDS facility can be reused. Users for the slag will likely be limited to those who are located in the vicinity of the plant. The economic viability of reuse decreases with increases in distance to market and hence increased transportation costs. If market conditions change so that commercial reuse ceases, the waste management techniques for the ash will need to be altered immediately. Long-term contracts with users may lessen the potential for market interruptions.

4.4.2.4.1 Control Function 2 - Treatment

The functional techniques applicable to treatment of the slag are fixation and encapsulation. Treatment of the slag may be appropriate if future data indicate that significant concentrations of trace metals were present in leachate from the slag ash. Currently available leach data indicate the trace metals in the leachate are low, and should not be significantly different from conventional coal boiler ash.

The performance of treating this stream is dependent upon the specific process (additive agent) used. Several lime- and cement-based fixation processes have been successfully applied to power plant ash, a material chemically not too dissimilar to the slag. Assuming that these fixation processes are applicable to treating this stream, the unit cost would be about \$10 - \$17 per Mg (100-103).

4.4.2.4.2 Control Function 3 - Disposal

The slag may be disposed of in two ways: in landfills or in impoundments.

4.4.2.4.2.1 Control Technique 1 - Landfill

In landfilling, gasifier slags are usually brought to the site by trucks, spread on the surface of land or previously placed wastes, and compacted. As discussed in Section 4.4.2.1.3.1, it is assumed that this stream will be disposed of with other solid waste streams in one landfill.

As discussed in Sections 4.4.1.3 and 4.4.2.1, annualized unit costs for landfilling the solid waste streams in a common landfill are estimated to be \$2.3 and \$4.1/Mg, respectively, for a non-lined and lined site. Based on these unit costs, the incremental annualized disposal cost for this stream would be \$1 and \$1.8 million dollars per year (not including administrative and hauling costs).

4.4.2.4.2.2 Control Technique 2 - Surface Impoundment

Surface impoundment is usually used for storage or disposal of wet ashes which are transported hydraulically to the impoundment in a fluid state. For storage impoundments, the ashes are dredged periodically and disposed in landfills. For disposal impoundments, the ashes are left in place and are covered to prevent erosion and infiltration of precipitations. The unit cost for surface impoundment is similar to the cost of landfill, assuming no excavation is required, although it is likely that larger volume of wastes will be handled due to the added water content.

4.4.2.5 Combined Boiler Bottom Ashes - Bottom Ashes from Steam and Power Generation Systems (Streams 704 and 710)

The characteristics of Streams 704 and 710 are the same, they consist essentially of coal ash; these two streams are thus likely combined for treatment or disposal.

The applicable control techniques for boiler bottom ash are similar to those for the Slag from Partial Oxidation Unit. Where landfill is the technique selected, the stream is assumed to be co-disposed with other solid waste streams in one common landfill.

As discussed in Sections 4.4.1.3 and 4.4.2.1 annualized unit costs for landfilling the solid waste streams in a common landfill are estimated to be \$2.3 and \$4.1/Mg, respectively, for an unlined and lined landfill. Based on these unit costs, the incremental annualized disposal cost attributable to this stream would be \$38,000 and \$68,000/yr for the base case, and \$47,000 and \$74,000/yr for the MFS case (not including administrative and hauling costs).

4.4.2.6 Combined Boiler Fly Ash - Fly Ash from Steam and Power Generation Systems (Streams 703 and 709)

The characteristics of Streams 703 and 709 are the same, they consist essentially of coal ash. The two streams are thus likely combined for treatment or disposal. The generation rate of these two streams is 63,000 Mg/yr in the base case, 77,000 Mg/yr in the MFS case.

Applicable control techniques for this stream are similar to those discussed for the Slag from Partial Oxidation stream. Available technology requires that gasifier ash and boiler bottom ash be quenched before any subsequent handling or disposal. Boiler fly ash, however, may be collected and handled dry (via a dry ESP or baghouse) or wet (via a wet ESP or scrubber). The choice of collection technology depends in part on site-specific disposal factors and also on factors specific to coal type. Some fly ashes tend to undergo fixation reactions when wetted, much as Portland cement does. When boiler fly ash is collected and handled entirely in the dry form, it can more readily be utilized as a resource.

Where landfill is the disposal technique selected, this stream is assumed to be co-disposed with other solid waste streams in one common landfill. As discussed in Sections 4.4.1.3 and 4.4.2.1 annualized unit costs for landfilling the solid waste streams in a common landfill are estimated to be \$2.3 and \$4.1/Mg, respectively, for an unlined and lined landfill. Based on these unit costs, the incremental annualized disposal cost attributable to this stream would be \$221,000 and \$394,000/yr for the base case and \$177,000 and \$316,000/yr for the MFS case (not including administrative and hauling costs).

4.4.2.7 Combined FGD Sludges - FGD Sludges from Steam and Power Generation Systems (Streams 705 and 711)

The characteristics of the two streams are the same. For lime/limestone FGD systems, they consist mainly of calcium sulfate and calcium sulfite, with some calcium carbonate. (If the FGD system incorporates forced oxidation, much of the sulfite will be converted to sulfate.) The two streams are likely combined for treatment or disposal. These streams are generated at a combined rate of 213,000 Mg/yr (base case) and 266,000 Mg/yr (MFS case).

4.4.2.7.1 Control Function 2 - Treatment

Where forced oxidation is not incorporated in the FGD system, FGD sludge typically contains 30 to 50 percent solids after thickening or filtration. It is not a good landfill material in this form because it is thixotropic. To rectify this problem, treatment by fixation may be practiced. (Where forced oxidation is employed, the solids might be dewatered to eliminate the thixotropic properties without fixation.)

4.4.2.7.1.1 Control Techniques 1 - Fixation

Several patented processes are available commercially for fixing FGD sludges. Typically these proprietary processes involve dewatering the sludge and combining the sludge with boiler fly ash and proprietary additives which promote pozzolanic reactions, resulting in a material less leachable, less permeable and structurally more suitable for landfill. Proprietary methods which have been successfully applied to fixing FGD sludges include Chemfix (addition of Portland cement and sodium silicate), Calcilox (calcined blast furnace slag and lime), IUCS - Poz - O - Tec (fly ash and lime under controlled temperature and moisture conditions, ICT (lime, betonite and cement), Research - Cottrell (sludge dewatering prior to fly ash admixing). Unit costs for these treatments range from $10 to $17 per metric ton of dry solids (100-103). Another treatment alternative practiced in many coal-fired power plants is mixing the FGD sludge with boiler bottom and fly ash before disposal. For coals that generate ashes that are alkaline, mixing the ash with the sludge will also initiate pozzolanic reactions.

4.4.2.7.2 Control Function 3 - Disposal

Disposal techniques applicable to FGD sludge include landfill and surface impoundment. Because of its low solids content and structural instability, FGD sludge may be treated by fixation prior to disposal. The use of surface impoundments will reduce the liquid content of disposal, but the dried solids are readily soluble when exposed to moisture after disposal, so proper surface impoundment closure will be essential.

As presented in Table 4-80, the flow rate of the FGD sludges stream is about 20% of the other inorganic ashes (such as power plant ash, gasifier ash and Flexicoking fines) combined. Assuming mixing the FGD sludges with these ashes generate a material suitable for landfill, the incremental annualized disposal costs for disposing the FGD sludges in a common landfill with all other solid wastes (at an annualized unit cost of $2.3/Mg for unlined landfill, $4.1/Mg for lined landfill) would be $490,000 and $870,000/yr for the base case, and $610,000 and $1,100,000/yr for the MFS case, depending on whether the landfill is unlined or lined. The basis and assumptions for these cost estimates have been presented in Sections 4.4.1.3.1, 4.4.2.1.3.1 and Appendix C.

Solid Waste Source Type 1
Dewatered Raw Water Treatment
Sludge
Control Functions 2 and 3

4.4.2.8 Stream 722 - Dewatered Raw Water Treatment Sludge

This stream - composed largely of calcium carbonate and water - is generated at a rate of 14,000 Mg/yr (base case) and 15,000 Mg/yr (MFS case). The applicable treatment and disposal techniques for this stream would be similar to those for the FGD sludge, except that the optimum fication process, the amount of chemical addition required and hence the treatment cost may differ. Where landfill is the technique selected, this stream is assumed to be co-disposed with other solid waste streams in one common landfill.

As discussed in Sections 4.4.1.3 and 4.4.2.1 annualized unit costs for landfilling the solid waste streams in a common landfill are estimated to be \$2.3 and \$4.1/Mg, respectively, for an unlined and lined landfill. Based on these unit costs, the incremental annualized disposal cost attributable to this stream would be \$32,000 and \$57,000/yr for the base case and \$35,000 and \$62,000/yr for the MFS case (not including administrative and hauling costs).

4.4.3 Source Type 2 - Recovered By-Products

This source type includes elemental sulfur recovered from the Bulk Sulfur Removal processes and collected dust from particulate control. The flow rate of these two streams is estimated to be, respectively, 29.5 Mg/hr (see Tables 4-18 and 4-19) and 0.16 to 2.6 Mg/hr (see Table 3-7). Among other by-products, recovered phenols are assumed to be burned as a plant fuel (and recovered ammonia is assumed to be marketed); hence the phenols are not expected to appear as solid waste.

4.4.3.1 Recovered Elemental Sulfur

Recovered elemental sulfur can be sold as by-product. However, the sulfur may be contaminated with carbonaceous impurities (from Claus plant) or vanadates, thiosulfates and thiocyanate salts (from Beavon Stretford tail gas treatment unit), making it non-marketable without further in-plant processing. If the sulfur stream cannot be sold, it can be disposed of in landfills.

There is a potential for elemental sulfur to be oxidized in a landfill environment, and such oxidation results in acid generation. Acidic leachate could solubilize trace elements from other wastes in the landfill. Hence it may be desirable to dispose waste elemental sulfur with alkaline wastes such as FGD sludges.

4.4.3.2 Collected Coal Dust from Particulate Control

This secondary waste stream consists primarily of coal dust collected throughout the coal preparation and liquefaction operations. The dust can be reused as feed to the gasifier or boiler, or can be disposed in landfills. When landfilled, spraying the dust with water may be required to reduce fugitive dust emission.

4.4.4 Source Type 3 - Organic Sludges

This source type includes one stream, namely, the biosludge from the biological treatment process. This is a secondary waste stream; the flow rate of this stream has been estimated to be 33,000 Mg/yr and 26,000 Mg/yr for the base and MFS case, respectively.

4.4.4.1 Stream 730 - Biosludge

As discussed in Section 4.3.1.1.4, the biosludge is assumed to consist of 80% water and 20% solid. The solid is mainly bacterial cell walls with some other organics. The potentially applicable control techniques are discussed below.

4.4.4.1.1 Control Function 1 - Treatment

Although no data are available on the composition of this waste at present, it is possible that some of the non-biodegradable toxic organics that might have been present in the raw process liquor (such as polycyclic organics and aromatic amines) will end up in the sludge through sorption. Thse organics can be destroyed by incineration.

4.4.4.1.1.1 Control Technique 1 - Incineration

Incineration of municipal and industrial biological oxidation sludges has been practiced for many years. The application of this treatment technique to this organic sludge could be expected to destroy greater than 99% of most organics and reduce the quantity of waste that requires ultimate disposal. Assuming the biosludge is 20% solids and 70% of the solids are combustible, the total waste quantity will be reduced by 94% after incineration.

Assuming a heating value of 5500 Kcal/kg (10,000 Btu/hr) for the biosludge, the heat input to the incinerator will be 2.3×10^7 kcal/hr (9.2×10^7 Btu/hr) and 1.8×10^7 kcal/hr (7.3×10^7 Btu/hr) for the base case and MFS case, respectively. Table 4-82 presents the estimated costs for incinerating the biosludge in a rotary kiln incinerator with energy recovery. The capital investments for the base case and MFS case are estimated to be, respectively, 16 to 20 million dollars and 15 to 18 million dollars (see Figure C6-2 in Appendix C6); or about 0.49% to 0.62% and 0.52% to 0.62% of the uncontrolled base plant costs. The total annualized costs for the two cases are 9.1 and 7.5 million dollars per year, respectively (see Figure C6-4); or about 0.95% and 0.86% of

TABLE 4 - 82. SUMMARY OF ESTIMATED COSTS FOR TREATMENT/ DISPOSAL OF BIOSLUDGE

Item	Treatment/Disposal			
	Incineration		Land Treatment	
	Base Case	MTS Case	Base Case	MFS Case
Total Capital Investment,* $\$10^6$	16 to 20	15 to 18	0.76	0.67
Total Annyalized Cost,** $\$10^6$	9.1	7.5	0.24	0.22
Unit Capital Investment, $\$10^6$/Mg/hr	3.8 to 4.7	4.5 to 5.3	0.18	0.20
Unit Annualized Cost, \$/Mg	280	290	7.4	8.5
% Base Plant Capital Investment,	0.49 to 0.62	0.52 to 0.62	0.02	0.02
% Base Plant Annualized Cost,	0.95	0.86	0.03	0.03

* Capital investments for incineration obtained from Figure C 6-2 (Appendix C6) for rotary kiln incinerators with capacities of 9.2 x 10^7 Btu/hr and 7.3 x 10^7 Btu/hr.

** Total annualized costs for incineration obtained from Figure C6-4.

the uncontrolled plant costs. See Appendix C6 for a more detailed description of the bases and assumptions for these cost estimates.

Two secondary waste streams are generated by this process, namely, a flue gas stream and a solid residue stream. It is not possible to estimate the characteristics of the flue gas, but the cost for controlling this is included in the incinerator cost estimates presented in Table 4-82. The incinerator is assumed to be equipped with a scrubber for particulate control. The scrubber blowdown would be routed to the wastewater treatment facility for treatment, if needed.

The flow rates of the residue stream are estimated to be 6 Mg/day and 4.8 Mg/day for the base case and MFS case, respectively. Assuming a 99.9% destruction of organics (see Appendix B-7 and Appendix C-6 for basis and assumptions for this estimate) the residue would contain about 0.33% organics, the remainder consisting of inert materials. Most of the trace metals originally present in the biosludge would be accumulated in the residue. Applicable additional treatment/disposal techniques include fixation/encapsulation and landfill.

4.4.4.1.2 Control Function 2 - Disposal

Biological treatment sludges may be disposed of in landfills, surface impoundment or by land treatment. Landfill and surface impoundment have been discussed in the previous sections. The following is a brief description of land treatment of biosludge.

4.4.4.1.2.1 Control Technique 1 - Land Treatment

In land treatment, biological treatment sludge may or may not require dewatering prior to applying to the land. Depending on the physical state, or the degree of dewatering performed, the sludges are transported to the land treatment site either by truck or hydraulic means. The sludges on land are spread with bulldozers, loaders, graders, or box spreaders. The site is generally subdivided into several plots which are treated in sequence. After waste application and evaporation of any associated water, the plot is plowed periodically until the waste has been decomposed. Chemicals such as nitrogen, phosphorus, and potassium may be added periodically as nutrients, and neutralizing agents (e.g., lime) may be added to maintain the proper pH level (7 to 9) (112,113).

The estimated costs for land treating the biosludges are summarized in Table 4-82. The capital investments are estimated to be 0.76 and 0.67 million dollars for the base case and MFS case, respectively. These costs are calculated from installed equipment costs using the the procedures summarized in in Table 4-1. The installed equipment costs include land preparation cost ($0.52/m^2), waste spreading equipment costs ($160,000) and monitoring well costs ($25,000). No land cost was included. It is estimated that about 430,000 m^2 and 340,000 m^2 of land are required for the base case and MFS case. Assuming a unit land cost of $5,000/10,000 per m^2, this would increase the total capital investment by more than 25%. The total annualized costs for the base and MFS case are, respectively, 0.24 and 0.22 million dollars per year. These costs are estimated using the unit costs presented in Table 4-2 and assuming a unit fuel requirement of 36,000 Kcal per Mg of waste landtreated and a labor requirement of 4,000 hrs/yr. No transporting nor other costs such as administrative, closure/post-closure and reliability costs are included in the annualized cost estimates. Depending on the distances, the transportation cost may more than double the annualized unit costs presented (at $2 and $4/Mg for a round-trip distance of 5 and 15 km, respectively).

4.4.5 Source Type 4 - Spent Catalysts

This source type includes 11 streams for the base case and 7 streams for the MFS case. Some streams such as Spent Sulfur Guard and Spent Drying Agent are not truly spent catalysts. They are included in this source type because (1) all of the streams in this source type are all small volume, intermittent streams, and (2) applicable controls are similar for these streams. The flow rates of streams that fall in this source type are summarized in Table 4-83. Although the flow rates are presented in Mg/yr basis, most streams are expected to be generated (removed) once every two or three years.

TABLE 4-83. LIST OF SPENT CATALYSTS WASTE STREAMS

Stream No.	Stream Description	Flow Rate, Mg/yr Base Case	Flow Rate, Mg/yr MFS Case
Stream 204 -	spent solvent hydrogenation catalyst (Ni-Mo)	821	821
Stream 404 -	spent hydrotreater catalyst from cryogenic hydrogen recovery (Ni-Mo)	68.5	37.2
Stream 405 -	spent drying agents from cryogenic hydrogen recovery (zeolite)	263	172
Stream 433 -	spent sulfur guard from hydrogen generation (ZnO)	271	0
Stream 435 -	spent reformer catalyst from hydrogen generation (Ni-U)	181	0
Stream 436 -	spent shift catalyst from hydrogen generation (iron oxide)	321	0
Stream 439 -	spent methanation catalyst from hydrogen generation (NiO)	115	0
Stream 453 -	spent drying agents from ammonia synthesis (alumina)	172	0
Stream 454 -	spent ammonia synthesis catalyst (iron/ iron oxide)	166	0
Stream 444 -	spent high temperature shift catalyst from hydrogen generation (Co-Mo)	0	451
Stream 445 -	spent low temperature shift catalyst from hydrogen generation (Co-Mo)	0	254
Stream 517 -	spent Claus catalysts (bauxite)	41	47
Stream 518 -	spent hydrolysis catalyst (Co-Mo)	20	20

4.4.5.1 Combined Spent Catalyst Streams - All Streams Listed in Table 4-83

Due to the proprietary nature of most catalysts, there is little information available on the reuse and disposal techniques applicable to specific catalysts. Because of this, spent catalyst treatment, reuse/resource recovery, and disposal are discussed in general terms in the following sections, with only brief mention of specific techniques that are applicable to individual catalysts.

4.4.5.1.1 Control Function 1 - Reuse/Resource Recovery

Spent catalysts may be reused after reactivation by a contractor or the original vendor; also the metals making up the catalyst may be recovered for other uses. The economics of the required regeneration processes and the market value of the metals will determine whether recovery and reuse are possible. In practice, return of the spent catalyst to the vendor for reprocessing will likely be the approach of choice in many cases.

Because of the current tight cobalt supply and the high demand for this metal, it appears that the cobalt-based shift catalyst could be economically recovered. Increasing cobalt prices have fostered interest by catalyst manufacturers to develop improved methods to regenerate the catalyst, to recover the metal, and to search for other catalysts (mainly nickel-based) which can be used in place of the cobalt-based shift catalyst (114).

Regeneration of spent drying agent (Streams 433 and 453), shift catalyst (436), ammonia synthesis catalyst (454) and Claus catalyst, is expected to be economically unattractive because of the low market value of the base materials of these catalysts (see Table 4-82). Spent methanation catalyst (Stream 439) and ammonia synthesis catalysts (Stream 454), although deactivated as far as catalyst activity is concerned, may still have a large capacity for absorption of sulfur compounds and can be used as sulfur guard bed material.

4.4.5.1.2 Control Function 2 - Treatment

Spent catalysts may be chemically fixed or encapsulated before final disposal to prevent leaching of undesirable substances. When fixing these with cement-based technique, the weight of the fixed material may be twice its original weight (1), i.e., a 1:1 chemical/spent catalyst ratio may be needed. As discussed before, the performance and cost for this alternative can only be established after thorough treatability studies.

4.4.5.1.3 Control Function 3 - Disposal

If these streams were not reused/reclaimed (per Section 4.4.5.1.1), they would likely be idsposed of in landfill. Assuming the unit disposal costs presented in Section 4.4.2.1.3.1 also apply to these wastes (i.e., assuming that these wastes are disposed of in a common landfill along with all other plant solid wates), the total incremental annualized cost for disposing these spent catalysts would be \$5,600 and \$10,000 for the base case, and \$4,100 and \$7,400 for the MFS case, depending on whether the landfill is non-lined or lined. Leaching and other characterization data of these streams, which are not currently available, may indicate additional disposal control requirements which would affect the disposal costs.

4.4.6 Integrated Control Examples

There are 26 solid waste streams identified in the base case, 23 in the MFS case. As discussed in Section 4.4.1, there are three control functions potentially applicable to controlling these streams. This section presents examples of how streams may be combined for ultimate disposal, and how certain control techniques may be combined to control a particular stream. The approaches considered in this section are by no means inclusive or representative of necessary or sufficient control.

4.4.6.1 Integrated Control Example 1 - Landfilling of Combined Waste Source Types 1, 3 and 4 Streams

As presented in Sections 4.4.2, 4.4.4 and 4.4.5, there are a total of 24 solid waste streams in the base case and 21 in the MFS case that belong to the Solid Waste Source Types 1, 3 and 4. The total annual flow rates for these streams amount to about 1.6×10^6 Mg/yr for the base case, 1.5×10^6 Mg/yr for the MFS case. One potential approach in handling these streams is to dispose all these wastes in one common landfill. It should be re-emphasized that this approach is just an example. These wastes may well be disposed in more than one landfill with different control designs or be handled with more than one control technique. By considering the alternatives of all wastes being disposed in an unlined landfill, and all wastes being disposed in a lined landfill, the range of landfill costs in the PCTM should bracket the costs that might be encountered in practice.

In this example, the wastes are assumed to be delivered to the landfill site by truck, unloaded at the toe of the slope of the working face, and then spread on top of the slope. To improve the structural stability of the fill, the FGD sludge would be mixed with fly ash or gasifier ash before being spread in layers over the old material. The biosludge, which constitutes about two percent of the total waste, is expected to be covered by the inert inorganic ashes all the time. Thus potential odor and site structural stability problems caused by this material would be alleviated.

The estimated costs for this integrated example is summarized in Table 4-84. The capital investments, not including the land cost, for both the base case and MFS case are the same, namely 11 and 23 million dollars, depending on whether the landfill is non-lined or lined (see Figure 4-8). These corrspond to 0.35% to 0.87% of the uncontrolled base plant cost. The total annualized costs for the base case (3.7 and 6.6 million dollars per year, respectively, for the non-lined and lined landfill) are slightly higher than the corresponding MFS values (3.5 and 6.2 million dollars per year) although the annualized unit costs ($2.3/Mg and $4.1/Mg, respectively, for the non-lined landfill) are the same as for both cases. The difference in total annualized costs is due to the fact that there is slightly more waste generated in the base case than the MFS case; however, this difference (1.6 million Mg/yr vs 1.5 million Mg/yr) is not large enough to affect the annualized unit costs. As discussed in 4.4.2.1.3, the annualized costs presented do not include administrative and hauling costs. Including such costs may increase the annualized costs by three times or more.

TABLE 4-84 SUMMARY OF COST ESTIMATES FOR SOLID WASTE INTEGRATED CONTROL EXAMPLES 1 AND 2

	Example 1				Example 2					
	Base Case		MFS Case		Base Case			MFS Case		
	Lined Landfill	Nonlined Landfill	Lined Landfill	Nonlined Landfill	FBC Boiler	Nonlined Landfill	Total	FBC Boiler	Nonlined Landfill	Total
Capital Investment, $\$10^6$	23	11	23	11	11	6	17	65	3	6.8
Total Annualized Cost, $\$10^6$	6.6	3.7	6.2	3.5	2.5	1.6	4.1	1.6	0.87	2.5
Total Annualized Unit Cost, $/Mg	4.1	2.3	4.1	2.3	2.1	2.4	4.5	2.7	2.9	5.6
% Base Plant Capital Investment, %	0.78	0.35	0.87	0.38	0.35	0.19	0.54	0.23	0.1	0.33
% Base P,ant Annualized Cost, %	0.70	0.39	0.70	0.39	0.22	0.25	0.47	0.18	0.1	0.28

4.4.6.2 Integrated Control Example 2 - FBC Boiler + Landfill of Combined Flexicoking Wastes

In this control example (see Figure 4-9), the Combined Flexicoking Waste Stream (Streams 302, 303, 306) was assumed to be used as feed to a Fluidized Bed Combustor (FBC) boiler (see Section 4.4.2.1.1). The ashes from the boiler are disposed of in a landfill. Burning the Flexicoking wastes in the boiler will recover the carbon heating values of the wastes. Landfilling the ashes generated from the boiler is an established disposal technique. As presented in Table 4-80, the feed to the FBC Boiler is 1.2 million Mg/yr of wastes for the base case, and 0.59 million Mg/yr for the MFS case.

Table 4-84 summarizes the cost estimates for this control example. The FBC boiler costs have been presented before (in Table 4-81). The landfill costs presented in Table 4-83 are based on estimates (see Figure 4-8) for landfills with 6.5×10^5 Mg/yr and 3.0×10^5 Mg/yr capacities. These are the expected ash quantities generated from the FBC boiler for the base case and MFS case, respectively. Non-lined landfill is used as an example since the residuals are essentially inorganic coal ashes. Some site-specific factors (such as high groundwater table) may impose additional disposal control requirements (such as lining).

Total capital investments for these examples, as presented, are estimated to be 17 and 6.8 million dollars for the base case and MFS case, respectively. The total annualized unit costs are \$4.5 and \$5.6/Mg for the two cases. As discussed before, no steam credits, land cost, hauling cost and other administrative costs were included in the cost estimates. Including these values would change the costs for this example.

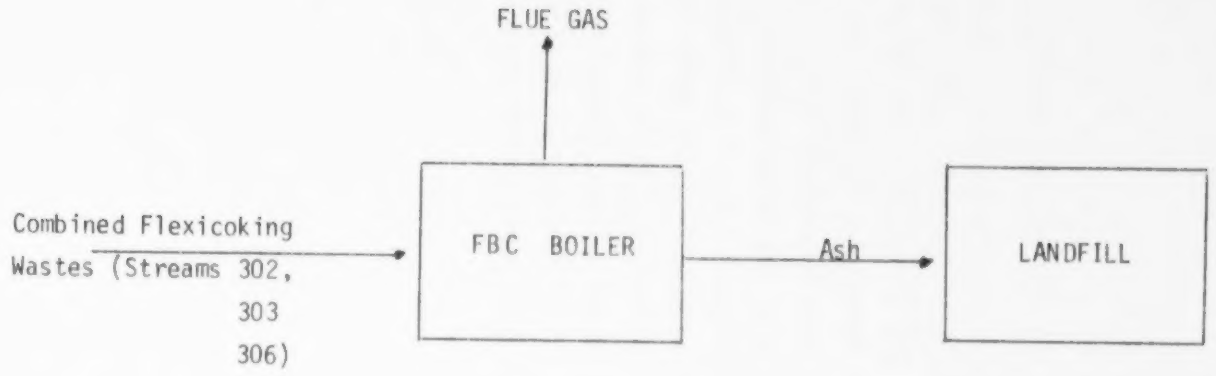
FLUE GAS
Combined Flexicoking
Wastes (Streams 302,
303
306)
FBC BOILER
Ash
LANDFILL

Figure 4.9. Solid waste control example 2: FBC boiler and landfill for Flexicoking wastes.

4.4.6.3 Integrated Control Example 3 - Incineration + Landfilling of Biosludge

Incinerating the biosludge will reduce the potential toxicity of and the volume of the waste that requires ultimate disposal. Assuming the biosludge is 20% solids and 70% of the solids are combustible, the total waste quantity will be reduced brom 100 Mg/day to 6 Mg/day for the base case, from 80 Mg/day to 4.8 Mg/day for the MFS case. Assuming annualized unit incineration costs of $280/Mg and $290/Mg for the two cases (see Table 4-82) and unit disposal costs of $4.1/Mg for a lined landfill (see 4.4.2.1.3.1) (i.e., assuming the incineration ash will be co-disposed with other solid waste in one lined landfill), total annualized costs for this example will be 8.9 and 5.4 million dollars per year for the two cases, respectively, or about 0.94% and 0.61% of the uncontrolled base plant costs. Since the landfill is assumed to accept other wastes, no incremental capital investment attributable to just the biosludge is estimated for this example.

5. DATA LIMITATIONS, GAPS AND RESEARCH NEEDS

Because of the inherent gaps and limitations which exist in the data base used to support this document, it is important for readers to understand the extent to which the performance and cost estimates presented here are supported by actual operating data, extrapolations from closely related applications or engineering calculations and/or judgements. The purpose of this section, therefore, is to convey to the reader a sense of the applicability and completeness of the data base. This information should contribute to a better understanding of how this document should be used by suggesting the confidence which can be placed in the uncontrolled discharge rates and the effectiveness of specific controls.

Waste streams which may be considered unique to EDS base case and MFS case designs have been emphasized in this section. These "unique" streams differ from analogous types of streams in related conventional industries (e.g., acid gases, organics-containing wastewater streams) in terms of detailed composition; these composition differences can result in differences in the performance and design of control equipment utilized. Unique streams are emphasized in Section 5 because the data limitations (concerning detailed composition, control performance/design) can be significant for these streams.

On the other hand, waste streams which are basically identical in composition to wastes generated in other industries (e.g., boiler flue gas, coal pile runoff, raw water treatment sludges and boiler ash) will experience control performances and designs which are identical to those employed for these streams in the related industries. These streams are not considered in this section. In general, much more is known about the composition of these conventional, "non-unique" streams compared to the streams unique to EDS; control performance/design is better understood through past experience in conventional industries, and data limitations are less severe.

Key data sources and the bases for characterization and control of unique gaseous, aqueous and solid waste streams are summarized respectively in Tables 5-1, 5-2 and 5-3 at the end of this section. Data gaps, limitations, and additional data needs for each waste stream are also summarized in the tables. For each waste stream, data sources and limitations are also summarized for some of the alternative individual control techniques that would potentially be applicable to that stream.

In general, characteristics of the uncontrolled unique waste streams were based on conceptual commercial design estimates provided by Exxon, and on data obtained by EPA and by Exxon on the 227 Mg/day (250 ton coal/day) EDS pilot plant at Baytown, Texas. The Exxon conceptual design estimates were derived using test data from small operating units, together with computer process synthesis. Wherever necessary, additional characteristics of the uncontrolled unique waste streams were estimated using data from other direct coal liquefaction processes (e.g., the Solvent Refined Coal-II process) and data from related industries such as petroleum refining and electric utilities. In all cases, the flow rates and compositions of all waste streams were based on an EDS commercial plant processing 1,134 Mg/hr (30,000 tons per stream day) of "as received" Illinois No. 6 coal in the liquefaction area. This EDS plant would produce about 9,580 m^3 (60,240 barrels) per stream day of fuel oil equivalent for the base case design, or 11,300 m^3 (71,080 barrels) per stream day of fuel oil equivalent for the MFS case design. These flow rates and compositions might, of course, vary in practice due to differences in design, coal type or plant operating conditions, and due to uncertainties in estimation. In a great number of cases, however, the uncertainties relating to the characteristics of waste streams would probably have limited impact on the selection of pollution control approach, and possibly limited impact on the expected performance of specific pollution control technology.

Since there are no existing EDS commercial plants, no direct operating experience is available to accurately indicate the detailed stream composition, or the performance or costs of applicable controls. For certain control systems (e.g., Claus bulk sulfur recovery and Wellman-Lord tail gas treatment) the existing data base from applications in related industries probably reflects gross pollutant removal efficiencies and associated costs in an EDS facility reasonably well. However, for control systems for which performance and cost are highly sensitive to individual components present in the waste stream (e.g., activated sludge and chemical oxidation), the limited data from conventional applications might be less indicative of performance/cost in an EDS plant. The limitations in the control performance data are twofold. First, the performance of control techniques in related applications is often known only in terms of major constituents, gross parameters or classes of substances. Little may be known about control performance for specific organics, trace elements or other trace constituents. Secondly, the performance characteristics of many controls are uncertain even for the major constituent and gross properties.

A data gap or limitation which exists for essentially all pollution control technologies relates to control reliability. Reliability data for many control techniques in conventional industries are limited. Differences between characteristics of waste streams from direct liquefaction and related industries only compounds the uncertainty of available reliability data.

TABLE 5-1. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR GASEOUS WASTE STREAMS AND AIR POLLUTION CONTROL TECHNOLOGIES

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Combined Acid Gas Stream		
Acid Gas from DEA Regenerator (Stream 508)		
Acid gas flow rate and composition, with the exception of COS, are based on estimates provided by EXXON. Concentration of COS is estimated by assuming that it is approximately equal to COS concentration in the acid gas from the DEA unit in the SRC-II commercial design. Reference: Section 3.3.3.3.1	No data on: 1) minor and trace components present in this stream or the sour gas streams entering the DEA scrubber, such as COS, CS_2, **mercaptans,** HCN, and heavy organics; 2) level of NH_3 present in this stream or sour gas streams entering the DEA scrubbers as a function of process operating conditions and **coal** type.	Distribution of sulfur species has **limited impact on selection of pollution con**trol approach or the sizing, performance, or cost of specific pollution control technologies employed. **However, the types and levels of other minor and trace components present need to be verified, especially if changes in process design or coal type are considered.**
Acid Gas from Sour Water Stripper/Ammonia Recovery (Stream 501)		
Acid gas flow rate and composition are based on estimates provided by EXXON. Reference: Section 4.2.1.1	No data on trace components present in the stream such as phenolics and other high molecular weight organics.	**The levels of trace components present need to be determined from demonstration or first commercial plants.**
Flash Gas from Partial Oxidation (Stream 440)		
Composition and flow rate estimates for this stream are based on Texaco gasifier pilot plant test data performed with SRC-II residue from Kentucky 9/14 coal. Reference: Section 3.3.4.2.1	No data on composition and flow rate estimates for this stream based on EDS residue from **Illinois No. 6 and other coals for a range** of EDS design/operating conditions.	Pilot plant tests in Texaco gasifier (or **other types of gasification units) with EDS vacuum bottoms slurry need to be performed to determine composition and flow rate of this flash gas, for a variety of coals and design/operating conditions.**
Selected Control Alternatives for Combined Acid Gas Stream		
Claus Process		
Performance and cost estimates are based on data from application in petroleum refineries, gas processing plants, and coke plants. However, the composition **for major species and flow rate of the combined acid gas stream from the EDS process are well within the range covered by these other applications.** References: Section 4.2.1, Appendix A-6.	The amount of NH_3 present in the combined acid gas is marginally high; any unburned NH_3 may result in the formation of solids such as ammonium hydrosulfide, ammonium polysulfides, ammonium carbonates or carbamates and/or ammonium sulfate causing solids deposition, solids plugging, and catalyst deactivation problems downstream.	Data needed to determine if the amount of NH_3 present in the combined acid gas is acceptable for conventional Claus plants, or whether alternate Claus designs or configurations are necessary to minimize **problems. Also, data are needed to verify performance of Claus plants in EDS demonstration or first commercial plants.**
Secondary Waste Stream		
Spent catalyst is the only secondary waste stream from the Claus process.	Generation rate and characteristics of spent Claus catalyst are not well known.	Determination of the leaching characteristics of spent Claus catalyst.
Beavon Sulfur Removal Process (BSRP)		
Performance and cost estimates are based on data for Claus tail gas treatment in petroleum refineries. References: Section 4.2.1, Appendix A-9	Uncertainties relating to the performance of the catalytic hydrogenation section under moderately high CO_2 concentrations (11-13%), especially the COS levels in the BSRP off-gas.	Determination of the COS levels in the effluent from the catalytic hydrogenator **under moderately high CO_2 concentration levels. Verify Beavon performance in EDS demonstration or first commercial plants.**

(Continued)

TABLE 5-1. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR GASEOUS WASTE STREAMS AND AIR POLLUTION CONTROL TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Secondary Waste Streams		
1) **Sour condensate**	**Characteristics of the condensate, particularly with respect to H_2S and NH_3 levels.**	**Determination of the characteristics of sour condensate is needed for the evaluation of control alternatives. Test control alternatives if needed.**
2) Stretford solution purge	Generation rate and characteristics of the purge stream are based on limited data and not well established. Resource recovery of this purge stream by reductive incineration has not been demonstrated on commercial scale.	**Acquisition of data from EDS demonstration or first commercial plants on characteristics of Stretford solution purge under different feed conditions. Demonstration of reductive incineration process on commercial scale.**
3) Spent Beavon catalyst	Generation rate and characteristics of spent Beavon catalyst are not well known.	Determination of the leaching characteristics of spent Beavon catalyst.
4) Stretford oxidizer vent gas	**Characterization data are not available for oxidizer vent gas. However, in EDS applications there probably would not be any gaseous components of environmental concern present in this vent gas.**	**Determination of the composition and flow rate of this vent gas in EDS demonstration or first commercial plants.**
SCOT Process		
Performance and cost estimates are based on data for Claus tail gas treatment in petroleum refineries.	No publicly available data on the performance of the catalytic hydrogenation section under moderately high CO_2 concentrations (15-18), especially with respect to the COS levels; no data on the removal efficiency for COS and CS_2 in the absorber; limited data on performance of DIPA versus MDEA in H_2S absorption and CO_2 co-absorption.	Determination of performance of SCOT catalytic hydrogenator and absorber under moderately high CO_2 concentrations; acquisition of data on DIPA versus MDEA performance as absorption **solution by tests in petroleum refineries or petrochemical plants. Verify SCOT performance in EDS demonstration or first commercial plants.**
Secondary Waste Streams		
1) Sour condensate	Characteristics of the sour condensate, particularly with respect to H_2S and NH_3 **levels.**	Determination of the characteristics of sour condensate is needed for the evaluation of control alternatives.
2) Spent SCOT catalyst	Generation rate and characteristics of spent SCOT catalyst are not well known.	Determination of the leaching characteristics of spent SCOT catalyst.
Wellman-Lord Process		
Performance and cost estimates are based on data for Claus tail gas treatment in petroleum refineries. References: Section 4.2.1, Appendix A-10.	Claus tail gas is incinerated prior to absorption in the Wellman-Lord process. **Once incinerated, there is no difference between the incinerated tail gas streams from EDS Claus plants and those from existing refinery Claus plants.**	**Verify Wellman-Lord performance in EDS demonstration or first commercial plants.**

(Continued)

TABLE 5-1. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR GASEOUS WASTE STREAMS AND AIR POLLUTION CONTROL TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Secondary Waste Streams		
1) **Acidic waste water**	Limited characterization data are available.	**Verify characteristics of this waste stream in EDS demonstration or first commercial plant.**
2) Thiosulfate/sulfate by-product purge	**Limited generation rate and characterization data are available.**	**Verify characteristics of this waste stream in EDS demonstration or first commercial plant.**
Thermal Incineration		
Performance and cost estimates are based on data for incineration of gaseous and liquid wastes. References: Section 4.2.1, Appendix A-16	None identified for incineration of off-gas from Claus tail gas treatment.	None identified for incineration of off-gas from Claus tail gas treatment.
Acid Gas from Acid Gas Removal Unit in Hydrogen Purification (Stream 428)		
Composition and flow rates for CO_2, H_2S, and H_2O are based on estimates provided by EXXON. Composition of other components are estimated using: 1) shift gas composition data from the Texaco gasifier in the processing of SRC-II residue and other Texaco gasification data; 2) performance characteristics of the Catacarb process. References: Section 3.3.4.2.1	Uncertainties relating to the composition of minor constituents, especially NH_3, COS, CO.	Determination of the levels of NH_3, COS and CO present in the shift gas from pilot plant tests. The concentration levels of these species may impact the selection of the pollution control approach and specific control technology.
Stretford Process		
Performance and cost estimates are based on data for feed gas containing a few percent CO_2 in gas processing and petroleum refinery applications. References: Section 4.2.1, Appendix A-7	Limited data on performance of the Stretford process for feed gas with high CO_2 levels (over 90 for Stream 428). There is some concern that CO_2 absorption by the Stretford solution would lower the pH and increase the bicarbonate/carbonate ratio, which in turn would lower the rate of H_2S absorption.	Acquisition of data on performance of the Stretford process for feed gas containing high levels of CO_2. The Stretford unit located in St. Elmo, Illinois and operating on stream consisting of 85 CO_2 and 15 H_2S would be a candidate test site.
Secondary Waste Streams		
1) **Stretford oxidizer vent gas**	**Generation rate and characteristics of oxidizer vent gas are not well known.**	**Determination of the generation rate, NH_3 level and trace contaminant levels of this vent gas, so that control alternatives can be evaluated.**
2) Stretford solution purge	No data on the generation rate and characteristics of this purge stream for feed gas containing high CO_2 levels. Resource recovery of this purge stream by reductive incineration has not been demonstrated on commercial scale.	Acquisition of data on the generation rate and characteristics of the Stretford solution purge in plants operating with high CO_2 feed gas. Demonstration of the reductive incineration process on commercial scale.

(Continued)

TABLE 5-1. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR GASEOUS WASTE STREAMS AND AIR POLLUTION CONTROL TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Flexicoking Heater/Gasifier Sour Fuel Gas (Stream 304)		
Sour fuel gas flow rate and composition are based on estimates provided by EXXON. References: Section 3.3.4.1.1	**Uncertainties concerning the levels and distribution of sulfur species, and the flow rate of this fuel gas.**	**Verify characteristics of this fuel gas stream in EDS demonstration or first commercial plants.**
Stretford Process		
Performance and cost estimates are based on data for feed gas containing a few percent CO_2 in gas processing and petroleum refinery applications. References: Section 4.2.1, Appendix A-7	None identified. Feed gas contains only 11.6% CO_2.	**Verify Stretford performance in EDS demonstration or first commercial plants.**
Secondary Waste Streams		
1) Stretford oxidizer vent gas.	**Characterization data are not available. However, feed gas does not contain NH_3 or mercaptans and these pollutants would not be present in the vent gas.**	**Determination of the composition and flow rate of this vent gas in EDS demonstration or first commercial plant.**
2) Stretford solution purge	Limited data on generation rate and characteristics of the purge stream. Resource recovery of this purge stream by reductive incineration has not yet been demonstrated on commercial scale.	Acquisition of data on characteristics of Stretford solution purge under different feed conditions. Demonstration of reductive incineration process on commercial scale.
Slurry Drier Vent Gas (Stream 102)		
Slurry dryer vent gas flow rate and composition are based on estimates provided by EXXON. Reference: Section 3.3.2.1.1	No data on minor and trace constituents of this vent gas.	Determination of the NH_3, COS, and mercaptan level in the vent gas in EDS **demonstration or first commercial plant.**
Vacuum Fractionator Off-Gases (Stream 153/156)		
Flow rate and composition are based on estimates provided by EXXON. Reference: Section 3.3.3.1.1	Uncertainty relating to the level of H_2S in the gas stream. Also, no data on minor and trace constituents of this gas stream.	Better definition of H_2S level in the gas stream. Determination of the NH_3, COS and mercaptan levels **and nature of organics in EDS demonstration or first commercial plant.**
Stretford Process		
Performance and cost estimates are based on data for feed gas containing a few percent CO_2 in gas processing and petroleum refinery applications.	None identified.	**Verify Stretford performance in EDS demonstration or first commercial plant.**
Secondary Waste Streams		
1) Stretford oxidizer vent gas	**Characterization data are not available. However, feed gas streams are not presently known to contain NH_3 or mercaptans.**	Determination of the composition and flow rate of this vent gas in EDS demonstration or first commercial plant.

(Continued)

TABLE 5-1. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR GASEOUS WASTE STREAMS AND AIR POLLUTION CONTROL TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
2) Stretford solution purge	Limited data on generation rate and characteristics of the purge stream. Resource recovery of this purge stream by reductive incineration has not been demonstrated on commercial scale.	Acquisition of additional data on characteristics of Stretford solution purge. Demonstration of reductive incineration on commercial scale.
Transient Waste Gases		
Transient Waste Gas from Liquefaction Reactor (Stream 803)		
Estimated to be similar in composition and flow rate to the liquefaction separator sour gas. Emission frequency and duration are based on engineering assumptions. Reference: Section 3.3.2.1.1	No data on flow rate, composition, emission frequency or duration for this waste stream.	**Data on the nature and levels of organics are needed. However, it would be difficult to formulate a research program to acquire these data.**
Transient Waste Gas from Flexicoking (Stream 801)		
Estimated to be similar in composition and flow rate to the Flexicoker fractionator off-gas. Emission frequency and duration are based on engineering assumptions. Reference: Section 3.3.4.1.1	No data on flow rate, composition, emission frequency or duration for this waste stream.	**Data on the nature and levels of organics are needed. However, it would be difficult to formulate a research program to acquire these data.**
Transient Waste Gas from Partial Oxidation (Stream 802)		
Estimated to be similar in composition and flow rate to the quenched syngas from the Texaco gasifier. Emission frequency and duration are based on engineering assumptions. Reference: Section 3.3.4.2.1	No data on flow rate, composition, emission frequency or duration for this waste stream.	**Data on the nature and levels of organics are needed. However, it would be difficult to formulate a research program to acquire these data.**
Flaring		
Performance and cost estimates are based on data for flaring of light hydrocarbons. References: Section 4.2.1, Appendix A-15	Limited data on destruction efficiency for CO and light hydrocarbons such as natural gas, ethylene, and propane. No data on destruction efficiency for high molecular weight organics expected to be present in transient waste gases.	Acquisition of data on destruction efficiency for representative organic compounds, especially C_4^+ organics and refractory compounds such as polycyclic organic matter (POM). This is a general data need for flares.
Thermal Incineration		
Performance and cost estimates are based on data for incineration of gaseous and liquid wastes. References: Section 4.2.1, Appendix A-16	Limited data on destruction efficiency for high molecular weight organics.	Acquisition of data on destruction efficiency for high molecular weight organics. This is a general data need pertinent to the incineration of hazardous waste streams.

(Continued)

TABLE 5-1. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR GASEOUS WASTE STREAMS AND AIR POLLUTION CONTROL TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Sodium Throwaway Process		
Performance and cost estimates are based on data for treating boiler flue gases.	None identified.	None identified.
References: Sections 4.2.1, 4.2.2, Appendix A-20		
Regeneration/Decommissioning Off-Gas from High and Low Temperature Shift Catalysts (Streams 449/450)		
Off-gas composition and flow rate are based on estimates provided in permit applications for the ANR Synthetic Natural Gas plant. Emission frequency and duration are based on engineering estimates.	No data or flow rate, composition, emission frequency or duration for this off-gas.	Acquisition of data on characteristics of this off-gas during the regeneration/decommissioning of sour shift catalysts in petroleum refineries, petrochemical plants, or SNG plants.
Reference: Section 3.3.4.2.1		
Sodium Throwaway Process		
Performance and cost estimates are based on data for treating boiler flue gases.	Limited or no data on performance under highly variable flow rate and high SO_2 loading (13%) conditions.	Determination of the practicality of controlling this off-gas with the sodium throwaway process. Acquisition of data on the removal efficiency for SO_2 at high inlet concentrations.
References: Sections 4.2.1, 4.2.2, Appendix A-20		
Vent Gas from CO_2 Removal by Catacarb Process (Stream 426)		
CO_2 and H_2 composition and flow rates are based on estimates provided by EXXON. Composition of other components are based on data for the Catacarb process.	Uncertainty relating to hydrocarbon and CO concentrations in vent gas.	**Verify characteristics of this waste stream in EDS demonstration or first commercial plants.**
Reference: Section 3.3.4.2.1		
Thermal Incineration		
Performance and cost estimates are based on data for incineration of gaseous and liquid wastes.	None identified for this waste stream.	**Verify incinerator performance in EDS demonstration or first commercial plant.**
References: Section 4.2.1, Appendix A-16		
Catalytic Incineration		
Performance and cost estimates are based on data for gas streams from coating ovens, printing plants, and manufacture of organic chemicals.	**Data on performance of catalytic incineration for this waste stream or similar waste streams are not available.**	**Verify catalytic incinerator performance in EDS demonstration or first commercial plant.**
References: Section 4.2.1, Appendix A-17		

(Continued)

TABLE 5-1. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR GASEOUS WASTE STREAMS AND AIR POLLUTION CONTROL TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Catalyst Regeneration Off-Gases		
Regeneration/Decommissioning Off-Gas from the Reformer Catalyst (Stream 446)		
Off-gas generation rate and composition are based on typical catalyst regeneration off-gas values. Emission frequency and duration are based on engineering estimates. Reference: Section 3.3.4.2.1	No data on flow rate, composition, emission frequency or duration for this off-gas.	Acquisition of data on characteristics of this off-gas during the regeneration/decommissioning of reformer catalyst in petroleum refineries, petrochemical plants, or SNG plants, and EDS demonstration or first commercial plants.
Decommissioning Off-Gas from the Methanation Catalyst (Stream 448)		
Off-gas generation rate and composition are based on typical catalyst decommissioning procedures. Emission frequency and duration are based on engineering estimates. Reference: Section 3.3.4.2.1	No data on flow rate, composition, emission frequency or duration for this off-gas.	Acquisition of data on characteristics of this off-gas during the decommissioning of methanation catalyst in petroleum refineries, petrochamical plants, or SNG plants, and EDS demonstration or first commercial plants.
Selected Control Alternatives for Catalyst Regeneration Off-Gas		
Flaring		
Performance and cost estimates are based on data for flaring of light hydrocarbons. References: Sections 4.2.1, 4.2.3.3.1.1, Appendix A-15	Data on flare performance for these waste waste streams are not available.	There is a general need for data on destruction efficiency of hydrocarbons in flaring. Also, verify flare performance in EDS demonstration or first commercial plants.
Thermal Incineration		
Performance and cost estimates are based on data for incineration of gaseous and liquid wastes. References: Sections 4.2.1, 4.2.3.3.1.2, Appendix A-16	Data on gas incinerator performance for these waste streams are not available.	Verify incinerator performance in EDS demonstration or first commercial plants.
Fugitive Organic Emissions from Process Equipment		
Emission estimates for fugitive organic emissions from leaking process equipment were based on emission factors for these components for conventional petroleum refining process equipment. Component counts were based on estimates provided by EXXON. Reference: Section 3.3.7.1	Characterization data are not available for EDS liquids or their vapors, or on leak frequency/rate in commercial scale direct liquefaction plants.	Exact composition of vapors from EDS plant process equipment may differ from petroleum refinery process equipment; however, the control technology that are applicable are not likely to be affected by these characteristics. Since these emissions directly enter the atmosphere, additional information on specific constituents of the vapor may be desirable. Also, leak frequency and rate might differ (e.g., abrasive slurries) from refineries; emission factors need to be developed based upon data from initial direct liquefaction plants.

(Continued)

TABLE 5-1. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR GASEOUS WASTE STREAMS AND AIR POLLUTION CONTROL TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Leak Detection and Repair Methods		
Leak detection and repair methods have been successfully used to control fugitive organic emission from petroleum refining equipment. This technique was assumed to be applicable to EDS process equipment. Emission estimates were based upon emission factors developed from test data for petroleum refining process equipment. References: Section 4.2.5, Appendix A-18	Emission factors for "controlled" fugitive sources (employing leak detection/repair methods) could differ compared to the conventional refinery case.	Determination of emission factors for fugitive organics sources employing leak detection/repair methods in EDS demonstration or first commercial plants.
Equipment Specification		
Replacement of leaking equipment with leakless components has been successfully used for the control of fugitive organic emissions from petroleum refining equipment. This technique was assumed to be applicable to EDS process equipment. Emission estimates were based upon emission factors developed from test data for petroleum refining process equipment. References: Section 4.2.5, Appendix A-18	Emission factors for "controlled" fugitive sources (employing equipment specification) could differ compared to the conventional refinery case.	Determination of emission factors for fugitive organics sources employing equipment specification in EDS demonstration or first commercial plants.
Evaporative Emissions from Product and By-Product Storage (Stream 751)		
Emission estimates were based upon emission factors developed from test data for conventional petroleum liquid storage. The amount and types of liquids stored and types of storage vessels were based on estimates provided by EXXON. References: Section 3.3.5.5.1	Characterization data are not available for EDS liquids or their vapors.	Exact composition of vapors from EDS product and by-product storage may differ from those of conventional petroleum liquids: however, the control technologies that are applicable are not likely to be affected by these characteristics. Since these emissions directly enter the atmosphere, additional information on specific constitutents of the vapor may be desirable.
Secondary Seals on Floating Roof Tanks		
Since secondary seals are used to control evaporative emissions from floating roof tanks storing petroleum liquids, they were assumed to be applicable to storage tanks containing synthetic liquids. Emission estimates were based upon emission factors developed from test data for petroleum liquids storage. Reference: Section 4.2.5	The performance of secondary seals is not expected to be different for EDS liquid storage tanks.	Research needs relative to the characterization of the emissions and not to the controls.

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TABLE 5-1. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR GASEOUS WASTE STREAMS AND AIR POLLUTION CONTROL TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Internal Floaters on Fixed Roof Tanks		
Since internal floaters are used to control evaporative emissions from fixed roof tanks storing petroleum liquids, they were assumed to be applicable to storage tanks containing synthetic liquids. Emission estimates were based upon emission factors developed from test data for petroleum liquids storage. Reference: Section 4.2.5	The performance of internal floaters is not expected to be different for EDS liquid storage tanks.	Research needs relative to the characterization of the emissions and not to the controls.

TABLE 5-2. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR WASTEWATER STREAMS AND WATER POLLUTION CONTROL TECHNOLOGIES

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Stream A and other High Phenolic Streams Liquefaction Cold Separator Wastewater (Stream 106), Atmospheric Fractionator Overhead Drum Wastewater (Stream 152), Vacuum Fractionator Overhead Drum Wastewater (Stream 155), Solvent Hydrogenation Fractionator Overhead Drum Wastewater (Stream 252)		
The characteristics of these streams are based upon Exxon design estimates and analytical data on the specific streams from the EDS pilot plants. There are differences between the pilot data sets, and between the pilot data and earlier design estimates.	Limited data on the characteristics of these waste streams are available from EDS pilot plant. Confirming data on a full range of organics are needed to allow better estimate of control performances. Also, data on conventional water pollutants and dissolved gases are needed.	Collect confirming data on a full range of organics, conventional pollutants, dissolved gases in EDS demonstration or first commercial plants.
References: Section 3.3.2.2.2, 3.3.3.1.1, and 3.3.3.2.2		
Solvent Hydrogenation Cold Separator Wastewater (Stream 202), Flexicoking Fractionator Overhead Drum Wastewater (Stream 308), Flexicoking Recontacting Drum Waster (Stream 307)		
The phenols and dissolved gases concentrations in these streams are based upon Exxon design estimates.	Same as above, except data are only available from small operating units.	Same as above.
References: Sections 3.3.2.1.2, 3.3.3.2.2, and 3.3.4.1.2		
Slurry Drier Cold Separator Wastewater (Stream 103)		
The phenols and dissolved gases concentrations in this stream are based on Exxon design estimates.	Same as above.	Same as above.
Selected Control Techniques for Stream A		
Phenosolvan		
The performance and cost estimates are based upon published data on Lurgi gasification plants.	The performance has not been determined for EDS wastewaters.	Determine performance of Phenosolvan for wastewater streams from EDS demonstration or first commercial plants.
References: Section 4.4.1.1.2.1, Appendix B-5		
Secondary Waste Streams		
Recovered Phenol	Data on the characteristics of this stream are not available. The purity of this stream may affect its resale potentials.	Characterization of this stream should be performed in EDS demonstration or first commercial plants to evaluate applicable resource recovery alternatives.

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TABLE 5-2. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR WASTEWATER STREAMS AND WATER POLLUTION CONTROL TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Chem Pro		
The performance and cost estimates are based upon vendor quotes, and data from the by-product coking industry and solvent extraction process.	The performance has not been determined for EDS wastewaters.	Collect performance data on Chem Pro in EDS demonstration or first commercial plants.
References: Section 4.4.1.1.2.2, Appendix B-5		
Secondary Waste Stream		
Recovered Phenol	Data on the characteristics of this stream are not available. The purity of this stream may affect its resale potentials.	Characterization of this stream should be performed in EDS demonstration or first commercial plants to evaluate applicable resource recovery alternatives.
Resin Adsorption		
The performance and cost estimates are based upon published data on organic chemical industries.	The performance has not been determined for EDS wastewaters.	The performance and cost effectiveness of this process on high concentration organic waste streams should be determined. Research effort should be directed toward determining feasibility of the process by concentrating on the following areas of study: maximum period of operating cycle, phenol and organics leakage during loading as a function of residence time, and regeneration requirements. Also, collect performance data in EDS demonstration or first commercial plants, if resin absorption were to be applied.
References: Section 4.3.1.1.2.3, Appendix B-10		
Secondary Waste Stream		
Recovered Phenol	Data on the characteristics of this stream are not available. The purity of this stream may affect its resale potentials.	Characterization of this stream should be performed in EDS demonstration or first commercial plants to evaluate applicable resource recovery alternatives.

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TABLE 5-2. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR WASTEWATER STREAMS AND WATER POLLUTION CONTROL TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Wet Air Oxidation		
The performance of this process is based upon published data on high COD containing wastewaters. Costs are based on vendor quotes.	Performance has not been determined for EDS wastewaters. Also, traditionally this process has been applied to waste streams with much smaller sizes (about one-tenth or less) than the EDS waste streams.	In principle, most organics in the EDS wastewaters would be oxidized under the operating conditions of this process. Collect performance data in EDS demonstration or first commercial plants, if wet air oxidation were to be applied.
References: Section 4.3.1.1.2.4, Appendix B-6		
Phosam-W		
The performance and costs are based upon published literature values on coking industry, Lurgi gasification plants and oil shale operations.	Performance has not been determined for EDS wastewaters.	**Collect performance data in EDS demonstration or first commercial plants.**
References: Section 4.3.1.1.3.1, Appendix B-8		
Secondary Waste Streams		
1) Recovered Ammonia	Characteristics of impurities present are not known.	Further characterization of this stream to check impurities is needed; this stream will likely be sold as a by-product.
2) Stripper Overhead	Characteristics of major components (NH_3, H_2S, water vapor) are known. Data on minor and trace components are lacking.	Further characterization of this stream with respect to organics is needed. Also, verify characteristics of this stream in EDS demonstration or first commercial plants, if applicable.
Chevron WWT		
The performance and cost estimates are based upon published data on refinery operations.	Performance has not been determined for EDS wastewaters.	Collect performance data in EDS demonstration or first commercial plants.
References: Section 4.3.1.1.3.2, Appendix B-9		
Secondary Waste Streams		
1) Recovered Ammonia	Characteristics of impurities present are not known.	Further characterization of this stream to check impurities is needed; this stream will likely be sold as a by-product.
2) Stripper Overhead	Characteristics of major components (NH_3, H_2S, water vapor) are known. Data on minor and trace components are lacking.	Further characterization of this stream with respect to organics is needed. Also, verify characteristics of this stream in EDS demonstration or first commercial plants, if applicable.

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TABLE 5-2. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR WASTEWATER STREAMS AND WATER POLLUTION CONTROL TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data Research Needs
Biological Oxidation (Activated Sludge)		
Performance is based upon that typically realized in applications in various related industries. Actual performance will be specific to the exact characteristics of the wastewater. Costs are based upon values reported in the literature, appropriately scaled on the basis of system loading. References: Section 4.3.1.1.4.1, Appendix B-10	Performance has not been determined for the specific constituents in EDS wastewaters. Only by direct testing of the wastewater to be tested can the performance of an activated sludge or any other biological treatment system be determined with certainty.	Bench-and pilot studies should be performed to determine the degree of biodegradability of the EDS wastewater. Samples for testing may be obtained from the EDS pilot plant. Also, test performance in EDS demonstration or first commercial process, including organics speciation upstream and downstream.
Secondary Waste Stream		
Biosludge	The flow rate and characteristic of this stream are not known. These can only be determined by testing with the actual wastestream. Also, performance of disposal alternatives (incineration, land disposal) are not known.	Characterization of this stream should be part of the research studies for the biological treatment system, and part of demonstration or commercial plant performance testing. Also, must characterize performance of disposal alternatives for biosludge (including incineration and land disposal), through monitoring around techniques used in EDS demonstration or first commercial plants, and perhaps through use of sludge generated in these plants for experimental incineration/disposal tests.
Activated Carbon Absorption		
Removal of all pollutants are based on data from the refining and by-product coking industries. Actual removal efficiencies will depend on pH, molecule size, and structure of the organics present in the wastewater. Costs are based on data from the refining and by-product coking industries, extrapolated on the basis of COD loading. References: Section 4.3.1.1.5.1, Appendix B-15	Performance has not been determined for the specific constituents in the EDS wastewaters.	Laboratory or pilot scale tests are needed on the EDS wastewaters. While many of the individual chemical species, determining COD are known, the performance cannot be synthesized from performances typical for individual species. Performance is generally specific to the wastewater being treated and therefore must be determined on an individual basis. Test performance in EDS demonstration or first commercial plants, including organics speciation upstream and downstream.

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TABLE 5-2. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR WASTEWATER STREAMS AND WATER POLLUTION CONTROL TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Chemical Oxidation		
Removals are based on the chemical oxidation amenability of individual species on the wastewater. Costs are based on data available in the open literature.	No experience exists for performance on specific constituents in the EDS wastewater. Overall pH dependence and general reactor requirements are not known. For wastewaters where organics are present, removals are uncertain, and unknown chemical species not present in the influent are likely to be found in the effluent.	Testing of the EDS wastewaters on a scale larger than laboratory scale is needed to better assess treatability. Destruction of chemical species by chemical oxidation under carefully controlled conditions are not necessarily realized in a full scale process. Test performance in EDS demonstration or first commercial plants.
References: Section 4.3.1.1.5.2, Appendix B-16		
Incineration		
Destruction of organics is based on that realised in other industries. Costs are based on vendor quotes.	Performance has not been established for EDS wastewaters.	Test performance in EDS demonstration or first commercial plants for which this technique is applied, including performance on individual organics.
References: Section 4.3.1.1.5.3, Appendix B-17		
Cooling Tower Concentration		
All nonvolatile chemical species with no exception are assumed to be concentrated into the blowdown stream without losses due to volatilization or drift.	Performance has not been determined for the EDS wastewaters. Loss of volatile species and potential for corrosion are not known.	Volatilization and composition of drift should be measured in EDS demonstration or first commercial plants for which this technique is applied.
Reference: Section 4.3.1.1.6.1		
Secondary Waste Stream		
Cooling Tower Drift	The characteristics of this stream are not firmly known.	Same as above.
Vapor Compression Evaporator		
All chemical species are assumed to be concentrated into the blowdown stream without losses due to volatilization. Costs are based on vendor quotes.	Performance has not been determined for the EDS wastewaters. The concentration of individual chemical species that will be carried over into the evaporator overhead is not known. Extent of corrosion potential is not known.	Test performance in EDS demonstration or first commercial plants for which this technique is applied, including characterization of species in overhead stream and in blowdown.
References: Section 4.3.1.1.6.2, Appendix B-26		
Secondary Waste Stream		
Concentrated Brine	The characteristics of this stream are not known.	Determine characteristics of this waste stream in EDS demonstration or first commercial plants.

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TABLE 5-2. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR WASTEWATER STREAMS AND WATER POLLUTION CONTROL TECHNOLOGIES **(Continued)**

Data Source and Basis	Data Gaps/ Limitations	Additional Data/Reseach Needs
Deepwell Injection		
All chemical species are assumed to be contained by this disposal technique. The costs are based on typical well construction and maintenance costs reported in the literature.	Pretreatment requirements for the EDS wastewater has not been determined.	Monitor deep wells in EDS demonstration or first commercial plants for which this technique is applied.
References: Section 4.3.1.1.7.2, Appendix B-27		
Surface Impoundment		
All nonvolatile chemical species are assumed to be contained by this disposal technique. The costs are based on typical impoundment construction costs reported in the literature.	Long term performance of liners and compatibility of liners with the EDS wastewater **have** not been determined.	Long term column or "cell" studies should be performed to determine the compatibility and the performance of liners. Monitor disposal sites in EDS demonstration or first commercial plants.
References: 4.3.1.1.7.3, Appendix B-28		
Combined Stream B and Other Low -Phenolic Streams Flexicoking Heater Overdrum Wastewater (Stream 312), Knockout Drum Wastewater in H_2 Cryo Recovery (Stream 403), Blowdown and K.O. Drum Wastewater From H_2 Generation (Stream 430), Catacarb Overhead Receiver Wastewater In H_2 Generation (Stream 431), Knockout Drum Wastewater In Ammonia Synthesis (Stream 452), Ageous Ammonia From Ammonia Synthesis (Stream 451)		
The characteristics of these streams are based on Exxon design estimates.	Characteristics of these streams are based on small bench scale tests and computer simulation studies. Characteristics of these streams from pilot or commercial size plant may differ.	Determine characteristics of these waste streams in EDS demonstration or first commercial plants.
References: Section 3.3.4.1.1, 3.3.4.2.2		
Sour Water From Partial Oxidation Unit (Steam 441)		
The characteristics of this stream are based upon gasification of a western coal.	Using the residues from the EDS process a feed to the gasifier may generate a stream with different characteristics.	Characteristics of this stream should be determined to evaluate control requirements. Characterization might be performed if EDS residues are gasified in an experimental gasifier; or in EDS demonstration or first commercial plants.
Selected Control Techniques for Stream B		
Techniques applicable to this set of streams have already been covered above in connection with Stream A (except Stream B will not need phenol extraction).		

TABLE 5-3. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR SOLID WASTE STREAMS AND SOLID WASTE MANAGEMENT TECHNOLOGIES

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
High-Carbon Solids		
Flexicoking Gasifier/Heater Dry Fines (Stream 302)		
Flexicoking Gasifier/Heater Wet Fines (Stream 303)		
Flexicoking Heater Bed Coke (Stream 306)		
The flow rates of these streams are based upon Exxon design estimates. Leachate characteristics are based upon data from EDS pilot studies.	Limited characterization data are available.	Verify characteristics of these waste streams in EDS demonstration or first commercial plants.
Reference: Section 3.3.4.1.3.		
Slurry Dryer Solids (Stream 108)		
The flow rate of this stream is based upon Exxon design estimates.	Data on the characteristics of this stream are not available.	Determine characteristics of this waste stream in EDS demonstration or first commercial plants.
Reference: Section 3.3.2.1.3.		
Selected Control Techniques for High Carbon Solids		
Fluidized Bed Combustion		
The burning of these materials in fluidized bed combustion (FBC) boiler is based upon an approach under consideration for a proposed K-T gasification plant in Alabama, and an existing K-T plant in Modderfontein, South Africa.	(1) The viability of commercial scale FBC boiler using these materials as feed has not been proven. (2) Although there are cost data for coal-fired FBC units, these data may not apply to combustion of these materials.	Perform tests on waste materials acquired from EDS demonstration or first commercial plants.
Reference: Section 4.4.2.1.1.1.		
Secondary Waste Streams		
FBC Boiler Fly Ash	No leachability data are available.	The leachability data should be determined in conjunction with FBC tests above.
Spent FBC Boiler Bed Sorbent	The flow rate and the leachability of this material are not known.	The flow rate and the leachability of this material should be determined in conjunction with FBC tests above.

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TABLE 5-3. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR SOLID WASTE STREAMS AND SOLID WASTE MANAGEMENT TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Landfill		
Landfilling of the waste is based upon current practices in the electric utility industry and other industries.	The long term leachate generation rates and characteristics are not known, and the compatibility and long term performance of landfill liners have not been established.	Long term column or landfill cell studies are needed to characterize the leachate and to determine liner performance. Monitor landfill associated with EDS demonstration or first commercial plants.
Reference: Section 4.4.2.1.2.1.		
Surface Impoundment		
Disposal of waste in surface impoundments is based upon current practices in the electric utility industry and other industries.	Same as landfill.	Same as landfill.
Reference: Section 4.4.2.1.3.2.		
Fixation		
Fixation of these materials to improve their structural stability and leachability characteristics is based on current practices for FGD sludge and other industrial wastes.	Performance and cost for the application are not known.	Treatability studies with various chemical additives might be performed to determine the technical and economical feasibility of this approach, on materials generated using EDS pilot bottoms in experimental Flexicoker, or on materials from EDS demonstration or first commercial plants.
Reference: Section 4.4.2.1.2.1.		
Spent Catalysts		
Spent Hydrogenation Catalyst (Stream 204)		
Spent Hydrotreater Catalyst (Stream 404)		
Spent Drying Agent (Stream 405)		
Spent High Temperature Shift Catalyst (Stream 444)		
Spent Low Temperature Shift Catalyst (Stream 445)		
Spent Claus Catalyst (Stream 517)		
Spent Hydrolysis Catalyst (Stream 518)		
The flow rates of these streams are based upon Exxon design estimates.	Leachability data on these streams are not available.	Leachability of these materials, if these catalysts are to be disposed of, instead of recycled should be determined. Appropriate materials for testing would be available one or two years after the first EDS plant is built.
References: Sections 3.3.3.2.3, 3.3.4.2.3.		

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TABLE 5-3. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR SOLID WASTE STREAMS AND SOLID WASTE MANAGEMENT TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Reuse/Resource Recovery		
These materials can potentially be reused by reactivation by a contractor or the original vendor, or reclamation of the metals making up the catalyst. Reference. Section 4.4.5.1.1.	The available markets for reuse of these materials are uncertain.	The availability of market is highly dependent upon local conditions which can only be assessed on a site-by-site basis.
Fixation		
Fixation of this material to improve its structural stability and leachability characteristic is based on current practices for FGD sludge and other industrial wastes.	Performance and cost for the application are not known.	Treatability studies with various chemical additives should be performed to determine the technical and economical feasibility of this approach, using spent catalysts from applications in related industries such as petroleum refining.
Landfill		
Landfilling of the waste is based upon current practices in the electric utility industry and other industries. Reference: Section 4.4.2.1.2.1.	The long term leachate generation rates and characteristics are not known, and the compatibility and long term performance of landfill liners have not been established.	Long term column or landfill cell studies are needed to characterize the leachate and to determine liner performance. These studies could be performed using similar spent catalysts from related applications such as petroleum refining.
Partial Oxidation Slag (Stream 442)		
The flow rate of the slag is based upon Exxon design estimates. The leachate characteristics are based upon pilot gasification of SRC-II flash drum bottoms using Kentucky No. 9/14 coals. Reference: Section 3.3.4.2.3.	Slag leachability data are not available for residuals from EDS process.	Slag leachability data should be acquired using residues from pilot plant gasification of EDS vacuum bottoms.
Flexicoking Chunks/Agglomerates (Stream 313)		
The flow rate and leachability data are based upon Exxon pilot studies. Reference: Section 3.3.4.1.3.	Limited characterization data are available.	Verify characteristics of this waste stream in EDS demonstration or first commercial plants.

(Continued)

TABLE 5-3. DATA GAPS, LIMITATIONS, AND RESEARCH NEEDS FOR SOLID WASTE STREAMS AND SOLID WASTE MANAGEMENT TECHNOLOGIES (Continued)

Data Source and Basis	Data Gaps/Limitations	Additional Data/Research Needs
Reuse as Construction Material		
Potential alternatives for utilizing these materials are based upon current practices for ash from coal-fired power plants.	The available market for utilization of these wastes (agglomerated coal ash) is uncertain.	The availability of market is highly dependent upon local conditions which can only be assessed on a site-by-site basis.
Landfill		
Landfilling of the waste is based upon current practices in the electric utility industry and other industries. Reference: Section 4.4.2.4.2.1.	The long term leachate generation rates and characteristics are not known, and the compatibility and long term performance of landfill liners have not been established.	Long term column or landfill cell studies are needed to characterize the leachate and to determine liner performance. Monitor disposal in EDS demonstration or first commercial plants. Perform column/cell studies for samples from EDS demonstration or first commercial plants, or for samples from experimental Flexicoker.
Surface Impoundment		
Disposal of waste in surface impoundments is based upon current practices in the electric utility industry and other industries. Reference: Section 4.4.2.4.2.2.	Same as landfill.	Same as landfill.

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